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SUPPLY AND SERVICES CANADA

(DEPARTMENT OF ENERGY, MINES AND RESOURCES)

FINAL REPORT (INCLUDING TASK 3)

ON

STUDY OF HIGH TEMPERATURE GAS CLEAN-UP

EBASTEC LAVALIN INC.

TORONTO, ONTARIO, CANADA

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ORIGINATED: J.W. Staines

APPROVED: B.J. Scarlett

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1. SUMMARY

1.0 Purpose of Study

In recent years the increased cost of the clean fuels, natural gas and oil, have caused more interest in coal and its derivatives as alternative lower cost fuels.

In Canada it is becoming evident that the large coal reserves in Western Canada could support gasification and other synfuel processes that are economically attractive for future energy needs. It would also assist Canada in achieving energy self-sufficiency.

Efficient utilization of fuel gas from coal gasification depends on the cost of gas cleaning to remove sulphur compounds and particulates that cause equipment and environmental problems. The current gas clean-up (purification) technology requires the use of a cold scrubbing pocess to cool gas to temperatures between 40°C and 150°C. The heat removed constitutes about 15 percent of the total available heat in the coal and, although partly recaptured in waste heat boilers, is a costly waste of energy.

Research and development is being carried out on a high-reliability gas turbine combined cycle Development Program using coal derived gas and oil. The gas turbine is designed to receive hot gases, without cooling from the gasifier. The hot gases entering the gas turbine must have been treated in a high temperature gas purification process to remove sulphur and nitrogen bearing compounds and particulate matter that cause excessive corrosion or wear.

The cleaning of hot fuel gas produced by the following three gasifier systems is the main purpose of this study. Gasification systems of direct interest are:

a) Moving Bed - Lurgi (low gasification temperature)

b) Fluidized Bed - Westinghouse (medium gasification temperature)

c) Entrained Bed - Texaco (high gasificaion temperature)

A comparison of the three systems will include the effect of gasification conditions on the amount of sulphur compounds, NO_x, and particulates in the gas.

The study includes the following three main activities:

Task 1 - A brief review of low temperature gas clean-up systems.

Task 2 - A detailed assessment of the state of the art in the development and commercialization of high temperature gas clean-up systems.

Task 3 - Economic and process comparison of low and high temperature gas clean-up systems. It includes the study basis and description of reference plants.

Final Report - This will include Task 3 and draw conclusions on the state of the art of the high temperature clean-up systems and on Tasks 1 and 2.

2.0 Scope of Study

The terms of reference require the submission of final and complete report including the findings to date on the completion of Task 3. Task 3 is the study basis (Section 4), description of reference plants (Section 7), and process and economic comparisons (Section 8). Task 1 was completed in Report

No. E-8205 and covered a brief review of low temperature gas purification systems. In it, one example of a low temperature clean-up system is identified for use in Task 3 of the study. Task 2 was completed in Report No. E-8301 and was a detailed assessment of high temperature gas cleanup systems.

3.0 Overview

This final and third report completes the comparitive evaluation of high and low temperature gas cleaning for coal gasification-combined cycle power systems.

It is concluded from the comparison of the unit costs of power $/kWH between high and low temperature purification for each of the Lurgi, Westinghouse and Texaco gasification processes that:

a) the Lurgi Dry Bottom Gasifier has the greatest benefit by replacing low temperature purification with high temperature purification. This results in a reduction in power cost from 0.0982 to 0.0872 $/kWH.

b) the Westinghouse and Texaco gasifiers both show increased power costs by replacing low temperature purification with high temperature purification. The power costs for the Westinghouse increase from 0.0816 to 0.0977 $/kWH, and for the Texaco unit gasifier from 0.917 to 0.0977 $/kWH. It appears therefore, there is no economic advantage in proposing high temperature purification for these gasifiers.

Although power cost assumptions between gasifiers are of limited validity because of the many assumptions, the power costs for the Lurgi with high temperature purification, the Westinghouse with low temperature purification and Texaco with low temperature purification all lie within a 12 percent scatterband. This indicates that the cost of power for the Lurgi with high temperature purification is probably competitive against the Westinghouse and Texaco gasifiers with low temperature purification.

2. CONCLUSIONS AND RECOMMENDATIONS

2.1 General

The basis for the study is the evaluation of the loss of heat energy from established gas clean-up (purification) technologies which require a cold scrubbing process to cool gas to temperatures between 40°C and 150°C prior to purification. Application of a hot purification system that would replace the cold system may reduce the heat loss and provide reduced pollution and reduced capital costs.

A range of gasifier systems was selected, i.e. Lurgi, Westinghouse and Texaco, for investigation to determine the incentive to adopt hot gas purification, and how this may be related to the gasifier type.

Low Temperature Purification

To provide a basis for comparison of high and low temperature purification, the low temperature clean-up technology in use for the three gasifiers is reviewed in Section 5. The low temperature sulphur removal technology required in low temperature purification is reviewed in Section 5 to select a commercially available low temperature (below 150°C) technology suitable for the three different modes of coal gasification. The Selexol process was selected because it can achieve the desired level of sulphur removal. The process is sufficiently selective to produce a suitable off-gas for use in a Claus sulphur removal unit. Selexol's low vapour pressure and high flash point tend to make solvent storage less hazardous in comparison to methanol.

High Temperature Purification

An assessment of the high temperature gas purification required for the different types of gasifiers is given in Section 6. The purification system includes both high temperature particulate removal and high temperature sulphur removal.

Dependent upon the specific gasifier and operating conditions, the particulate loadings may be sufficiently high to require gas clean-up prior to sulphur removal and/or combustion turbine equipment to achieve acceptable life and performance. Alkali metals in different physical forms (vapour, solid) may also be present at levels that require reduction prior to gas turbine combustion. The review of high temperature particulate removal technology included cyclones, electrostatic precipitators, ceramic fiber filters, granular bed filters and novel technologies. It is recommended for the purposes of this study, a combination of cyclones and ceramic bed filters be installed for sulphur absorption vessel and combustion turbine equipment protection. The cyclones would provide a gross solids separation (10 microns) with ceramic fibre filters removing the finer particulate.

The principal objective of a high temperature sulphur removal system for coal gasification fueled combined cycle power generation is to reduce the sulphur components in the fuel gas to levels which are compatible with environmental emissions and combustion turbine equipment requirements without significant expenditure or loss of efficiency. Currently, most of the present efforts are in the research and development stages and have involved laboratory testing of various metal oxide sorbents. Zinc ferrite sorbents appear to be the most favourable, of the studied metal oxides, for further development and testing for possible commercial application.

Process Comparisons

The process comparisons of Section 8 provide the basis for determining the economic incentive, if any, for development of high temperature purification and if there was any difference in incentive for a particular gasifier type.

In Table 8.1-1, of Comparitive Thermal Efficiencies for low versus high temperature purification, the improvement for high temperature purification is 24.4 percent for the low temperature Lurgi gasification process, 5.9 percent

for the moderate temperature Westinghouse process and 7.7 percent for the high temperature Texaco process. This indicates that the Lurgi gasification process achieves the greatest thermal benefits from high temperature purification. The high temperature purification (Case MH) incorporates particulate removal (cyclones and ceramic filters) and desulphurization using zinc ferrite, and it is expected that the hydrocarbons including tars pass unchanged in vapour form through this equipment. The hydrocarbons are consequently converted to electricity at the combined cycle efficiency of over 45 percent.

In the low temperature Lurgi gasifier process (Case ML) the large quantities of steam used to prevent ash clinkering passes through the gasifier and is condensed in the gas quench operation necessary for removal of tar, phenols, oils and particulate. The recovered hydrocarbons are supplementary fired after the combustion turbine and eventually converted to electricity at the steam cycle efficiency of about 33 percent.

The underlying reasons for the high thermal benefits to the high temperature purification Lurgi gasifier process (Case MH) and small benefits to the other high temperature purification cases, Westinghouse (Case FH) and Texaco (Case EH), can be attributed to the thermal losses in the low temperature Lurgi gasifier process (Case ML).

The raw gases leaving the higher temperature gasifiers with low temperature purification (Westinghouse Case FL and Texaco Case EL) contain substantial amounts of steam, without tars and other high boiling hydrocarbons. The high level sensible heat in the gas is recovered in boilers before quenching to remove particulate matter. Therefore with these gasifiers the change in heat usefully recoverable that comes with high temperature purification (Westinghouse Case FH and Texaco Case EH) is not as great as with the Lurgi gasifier (Case MH).

Technical Risk

This study determines the projected change in the cost of producing electricity when the purification of the coal derived gases is accomplished at high temperature. The risk is that associated with the high temperature purification process and includes the high temperature sulphur removal (zinc ferrite bed) and particulate collection systems (ceramic fibre filter) which are only at the laboratory stage of development. The process conceptual design is therefore only order of magnitude and capital and operating costs based on it must be appraised as soft.

Environmental Impact

An assessment of the environmental impact of the six gasification plant technologies did not reveal any potential cause for air quality or ground water quality degradation. However, it is assumed that high NO_x emissions from the combined cycle gas turbine for the high temperature purification cases will be reduced by future new design combustors.

Economic Comparisons

The economic comparisons of the unit cost of power $/KWH (1982 prices) between high and low temperature purification for each of the Lurgi, Westinghouse and Texaco gasification processes indicates that:

a) The Lurgi Dry Bottom Gasifier has the greatest benefit by replacing low temperature purification with high temperature purification.

b) The Westinghouse and Texaco gasifiers both show increased power costs by replacing low temperature purification with high temperature purification. It appears there is no economic advantage in proposing high temperature purification for these two gasifiers.

Sensitivity Analysis

This illustrates the trends of levelized power cost for variation of parameters; plant capacity, capital cost, operating cost and inflation, when these are varied. Power cost increases with reduction in plant capacity from the base of 70 percent and is due to the effect of the fixed costs.

An increase in capital cost causes a relatively small increase in power cost and this can be attributed to the high ratio of operating costs to capital costs. The reverse effect occurs with operating cost when an increase in power cost causes a large increase in operating cost.

Inflation at a 6 percent compounded rate causes a 60 percent power cost increase over the life of the plant. A real increase of one percent in the cost of coal creates a 4.8 percent increase in the cost of power. The cost of power is therefore very sensitive to an increase in the inflation rate or in the cost of coal.

Conclusion

The process comparisons indicate that the Lurgi gasification process achieves the greatest thermal benefits from high temperature purification (MH) because of the thermal losses associated with tar removal in the low temperature purification (ML) are avoided.

The small increase in thermal efficiency for the higher temperature gasifiers (Westinghouse and Texaco) is due to the low temperature purification gas not containing tars. The high sensible heat in the gas can therefore be recovered in boilers before quenching.

The economic comparison of the unit cost of power between high and low temperature purification for each of the Lurgi, Westinghouse and Texaco gasification processes indicates that the Lurgi alone benefits from high temperature purification, and that increased power costs are associated with the Westinghouse and Texaco gasifiers.

The significant loss in thermal efficiency in the case of Texaco and to a lesser extent Westinghouse gasification processes is due to cooling the gas to the 600-700°C operating range of the high temperature desulphurization beds. The absorption capacity of the desulphurizing materials considered declines steeply above the stated operating temperature range. Additional research and development with the aim of increasing this temperature would appear worthy of serious consideration. Further development of the high temperature particulate removal systems would be required at higher temperatures to parallel this work.

3. DESCRIPTION OF STUDY

3.1 Objectives

Application of high temperature purification of coal derived fuel gas in combined cycle power generation may lead to increased energy efficiency, reduced environmental pollution and reduced capital costs. The high temperature gas purification process would replace the established water scrubbing processes for particulate matter removal and low-temperature sulphur removal processes.

The high temperature gas purification processes for removing sulphur compounds and particulate matter from the fuel gas would justify application provided that it improves energy production economics to levels significantly above current experience and satisfies environmental pollution constraints. The objectives of this study are to identify the potential advantages of high temperature purification technology for combined cycle power generation which may be economically justified and to review and comment on developing high temperature purification technologies.

3.2 Approach

It is recognized that the amount and direction of change in the cost of electric power produced in a coal gasification combined cycle (CGCC) that is attributable to use of high rather than low temperature purification of the fuel gas can be influenced by the gasification temperature level and mode of gasification. To quantify this, three gasification technologies were selected each typifying gasification at a different level of temperature. Low temperature gasification is typified by a Lurgi dry bottom moving bed gasifier; fluidized bed gasification in a Westinghouse gasifier represents medium temperature gasification; and for high temperature, the entrained flow gasification as practiced by Texaco is used. For each gasification technology, two alternate combined cycle plant designs were sufficiently developed for hot and cold purification fuel gas to permit calculation of the thermal efficiency and

operating requirements as well as estimation of their capital investment needs. One alternate for each of the gasification technologies utilizes conventional present day technology for purification of the raw gas at low temperature (about 35°C) while the second alternate uses a conceptual design of facilities for particulate and sulphur removal at high temperature (about 650°C). All three gasification technologies use the same selected process for low temperature sulphur removal and the same conceptual high temperature purification means.

Financial analysis using the calculated operating requirements and the estimated investment costs provides the comparative costs of electricity produced using the high and low temperature purification for a given gasification technology.

The two alternates evaluated for each gasification technology incorporate, to the greatest practical extent, the same process design philosophy so as to minimize the effect that factors other than the temperature level of purification have on the comparisons made. Some of these factors include steam pressures and temperatures, temperature approaches for heat recovery, and mechanical efficiencies.

The purification means, both and hot used with all three gasification technologies are selected because of their process advantages, but may not be the best possible or optimum to be used with any of the gasification technologies. The purification means incorporated in the Study, however, are suitable for use with any of the gasification technologies and are not prejudicial to any of them. No attempt is made as part of this Study to optimize the gasification technologies and the purification means.

The same type, Illinois #6, is used for all three gasification technologies and is considered eminently suitable for use in any of them. Use of a common coal throughout the Study means that the degree of relative change in the cost of electric power which is evidenced between hot and cold purification for a

temperature level of gasification can validly be compared to that evidenced for other gasification temperature levels.

The approach or methodology used for this Study result in approximate values for power cost and thermal efficiency for combined cycle power systems. This is because of the different stages of development of the gasifiers and the early stage of development of high temperature purification systems. Therefore, comparisons made between the gasification technologies using either hot or cold purification are of limited validity. Caution must be used in selecting the lowest cost of power generated in a CGCC plant from any of the six combinations of gasification-purification technologies.

However, for a CGCC plant using a given gasification technology, the approach does result in the incremental difference in the cost of electric power that can be expected and attributable to the use of high temperature purification instead of present day low temperature purification.

4. STUDY BASIS

Continuing attention is being paid to coal as a potential source of gaseous fuel to fire combustion gas turbines operating in combined cycle mode for electric power generation. Such operation, depending upon needs at the point of application, could also incorporate cogneration of steam for process and/or heating purposes.

Central to the design of an optimum coal gasification energy conversion system is the selection of the gasifier. The type of gasifier used is a major factor in the complexity of the overall system and consequently of the capital and operating costs. The gasifier and its accompanying cleanup system has a significant effect on overall plant thermal efficiency (coal to kilojoules).

If it is assumed that the steam cycle or combined gas/steam plant cycle is constant the thermal efficiency of the gasifier/cleanup system has a significant effect on overall plant efficiency. Considering a contemporary combined cycle plant of 50 percent efficiency the gasifier/cleanup system must be 75 percent efficient to give the same overall plant efficiency as the 37 percent of a modern base load steam plant with stack gas cleanup.

The expected performance of the three gasifiers described in this section lies in the range of 76 to 78 percent gas efficiency. It becomes obvious that without the recovery of sensible heat in the product gas only a few gasifiers have the gas efficiency to provide an integrated system competitive with existing large steam plants. Therefore, the recovery of product gas heat makes plant integration an economic necessity. It is anticipated that high temperature gas cleanup may result in increased product gas heat recovery.

A broad spectrum of coal gasification processes have been proposed and are available for gasifying coal to produce a gaseous fuel. In all cases the conversion of coal to produce a gaseous mixture containing combustibles requires elevated temperatures, above 750°C, and the aid of steam as a gasify-

ing medium. The crude gas leaving the gasifier proper at a high temperature consists mainly of CO, H_2, CO_2 and CH_4 but may also contain considerable impurities and trace elements. The considerable heat required to effect the conversion of the coal is usually provided by simultaneously burning some of the coal, with the aid of oxygen directly in the gasifier. If air is used directly to supply the oxygen needs, the crude gas produced would contain considerable nitrogen (about 50%).

The characteristics of the gas leaving the gasifier itself is greatly influenced by the coal feedstock, the gasifier design, and its mode of operation. Coal is a very complex substance highly variable in composition even when taken from the same seam, and contains many inorganic and metallic impurities. Nearly any element found in the coal may be expected to be present in some form and to some extent in the crude gas, but this depends largely on the gasification temperature and gasifier design.

4.1 Coal Analysis

For purposes of this study a coal having the following characteristics and referred to as an Illinois #6 coal has been selected:

Proximate Analysis

Moisture	4.2%
Ash	9.6
Fixed Carbon	52.0
Volatile Matter	34.2
	100.0

Ultimate Analysis - DAF Basis

Carbon	77.26%
Hydrogen	5.92
Oxygen	11.14
Nitrogen	1.39
Sulphur	4.29

Heating Value - As Received

Higher Heating Value (HHV)	28,459 kJ/kg (12,235 Btu/lb)
Net Heating Value (LHV)	27,235 kJ/kg (11,709 Btu/lb)

Size - As Received

Moving Bed Cases (ML and MH)	Sized 38 mm x 6.4 mm
Fluid Bed (FL and FH) and Entrained Flow (EL and EH) Cases	Sized 38 mm x 0

This coal which is ranked as high volatile B bimuminous coal by ASTM D 338, was chosen because it would not unduly prejudice any of the gasifier technologies preselected for this study. It is a type of coal which has been tested and found suitable for use in all the gasifiers of this study. As a result of it being widely studied and evaluated there is available substantial technical and economic information in the public domain.

4.2 Raw Gas Analysis

The gasifier configuration and its mode of operation to a large degree determines the composition of the gas produced from a given coal and the variety of impurities the gas contains. Of principal importance are the operating temperature (or temperature range), the contact time, and the method of contact or degree of countercurrentcy between the coal and oxidants.

Moving bed gasifiers (sometimes referred to as fixed bed because of the relatively slow movement of the coal/ash particles downward through the gasifier) provide a long coal solids residence time which while promoting a higher degree of carbon conversion does result in relatively low gasifier throughputs. The gases flow countercurrent to the solids, which are supported by a grate, for maximum thermal efficiency. In a fluidized-bed gasifier the gases move through the mass of coal particles at a velocity sufficiently high to support their weight but not high enough to carry excessive solids out of the gasi-

fier. This gasifier exhibits uniform temperatures throughout the fluidized beds and good contact between solids and gas. Entrained gasifiers use sufficiently high gas velocities to entrain or carry the solid particles co-currently in a gas-solid suspension. High gasification rates or throughputs can be achieved in the entrained gasifiers.

Hydrogen, carbon monoxide, carbon dioxide and water vapour are major constituents in the crude gas from most gasifiers. These exist is proportions dictated by the water-gas shift reaction equilibrium and are, therefore, dependent on the temperature and quantity of steam used. Gasification systems operating at lower temperatures tend to produce a higher level of methane and to a much lower extent other light hydrocarbons. If air is used to provide the oxygen needs of the gasifier, the crude gas would also contain considerable nitrogen.

Dependent on the degree of countercurrent contact the lower temperature gasifiers will tend to produce, primarily by devolatilization of the incoming coal, condensible hydrocarbons which may range from very light benzene-toluene-xylene mixtures or naphthas through coal tars. Although the oxygen of the coal is largely converted to carbon oxides and water significant amounts of phenols are produced in countercurrent gasifiers at temperatures below about 650°C in the upper or coolest stage of the bed.

Essentially, all the sulphur contained in the coal appears in the crude gas as hydrogen sulphide and carbonyl sulphide. The relative proportion is largely temperature dependent with higher temperatures favouring the presence of COS. The ratio existing between these principal sulphur compounds is also dependent on the steam to coal ratio used. Some sulphur tends to appear as carbon disulphide, mercaptans and other organic sulphur compounds in those processes that tend to produce tars and oils (i.e. low temperature gasifiers).

Fixed nitrogen within the feed coal can react with hydrogen during gasification to form ammonia, which at high temperatures can further react with carbon to form cyanides:

$$2N + 3H_2 \longrightarrow 2NH_3$$
$$NH_3 + C \longrightarrow HCN + H_2$$

Low temperature countercurrent gasifiers tend to have at least trace amounts of nitrogeneous compounds, such as pyridines amines, etc, present in the crude gases. These probably are a result of dissociation of the complex high-molecular weight nitrogen-containing organic compounds originally present in the feed coal. Because ammonia tends to decompose to nitrogen and hydrogen at high temperatures, high-temperature gasifiers are expected to have very little ammonia in the crude gas.

The crude gas from the gasifiers may contain not only solid particles, liquid droplets, and sticky or plastic phase materials but also impurities in the vapour phase. Coal contains many inorganic metallic impurities or minerals which makeup the ash. They appear predominantly in the ash or slag withdrawn from the gasifiers but some appear as impurities and particulate matter with the crude gas. Dependent on the characteristics of the coal and the operating mode and conditions of the gasifier some unreacted and partially reacted coal may also appear as soot or char in the crude gas.

Unfortunately the information needed to characterize and quantify the minor components, the impurities, and the contaminants in the coal-derived gases is not presently available. Further, some of the information available is questionable. This is particularly true with respect to the hot effluent stream leaving the gasifiers themselves. At that location the obtaining of a true representative composite sample is extremely difficult and its handling so as to obtain analytical results, truly representative of its condition when hot,

is impractical. The hot effluent from the gasifier is highly unstable as to its chemistry and physical state.

Much of the information available for quantifying the crude gas leaving the gasifiers is inferential in nature. The bulk of the information has been obtained with gasification systems incorporating means for scrubbing and/or cooling the gas before its discharge to the atmosphere or its end use. Under these circumstances the need to or the incentive to characterize the crude gas from the gasifier itself is non-existent as long as the product gas and other major streams from the system meets the overall requirements.

With these limitations in mind, typical characteristics of the hot crude effluent streams from the three type gasifiers being considered in this study are presented in Table 4-1. The quantities used to quantify the characteristics should not be considered absolute but considered an attempt to show the relative effect of the mode of gasification on the effluent stream characteristics.

TABLE 4-1

GASIFIER CHARACTERISTICS

	Moving Bed	Fluidized Bed	Entrained Bed
Gasifier Technology used for study	Lurgi (dry bottom)	Westinghouse	Texaco
Temperature classification	Low (2)	Moderate (3)	High (4)
Gasification Reaction Temperature	700 to 850°C	950 to 1050°C	1250 to 1450°C
Type Flow	Counter-current	-	Co-current
Residence time for coal	Long (up to 60 min)	Moderate (about 15 min)	Short (about (5 seconds)
Oxygen to Coal Ratio used in Study ([5])	0.54	0.70	0.95
Water to Coal Ratio used in Study ([6])	2.63	0.78	0.65
Approximate Gas Analysis dry basis: CH_4 and C_2's	9.5	8.2	0.1
H_2	43.3	32.8	36.0
CO	15.5	42.5	53.0
CO_2	31.7	16.5	10.9
	100.0	100.0	100.0
Steam to Gas Ratio by Volume	0.505	0.213	0.179
H_2 to CO Ratio by Volume	2.8	0.77	1.0
(CO_2 x H_2) to (CO x H_2O) ratio	0.845	0.539	0.391
H_2S to COS Ratio in Gas, molar	Approx. 23	NA	Approx. 17
Particulate Material, kg/m^3	Approx. 0.2	Approx. 0.4([7])	Approx. 0.7([8])
Alkali Concentration, ppm by weight: Sodium	0.028-0.035	19-23	87-12.3
Potassium	0.013-0.016	4.3-12	4.5-6.4
Organic Compounds (i.e, naphthas, tars, oils, phenols, etc.)	Total quantity equal to 8 to 10 percent of coal carbon content	Trace amounts	None

Notes for Table 4-1:

1. NA - not available.

2. Maximum temperature in the "dry bottom" variant of the moving bed type gasifier occurs in the combustion zone and is limited to avoid stickiness or agglomeration.

3. Sufficiently high to cause some agglomeration.

4. Sufficiently high for ash to be molten.

5. Kilograms oxygen per kilogram coal on a moisture, ash free basis.

6. The kilograms of water per kilogram of moisture, ash free coal includes the moisture associated with the coal and the water and/or steam added. For the moving and fluidized bed gasifiers the addition is as steam while for the entrained bed gasifiers the addition is as water to produce a feed slurry concentration of 66.5 percent solids.

7. Particulate loading was developed from published reports and verbal communication with Westinghouse. Average particulate size is 2500 micrometers exit gasifier itself. About 96 percent of total loading is removed in cyclones associated with gasifier and operating hot. Of the minus micrometer particles leaving the secondary cyclone about 50 percent are plus 10% and 98 percent plus 1.5 micrometers.

8. After flow reversal in the radiant boiler located below gasifer the particulate loading is about 0.1 kg per cubic metre.

9. Loadings were calculated using thermodynamic model to rank gasification systems on the basis of released volatile alkali concentrations at pro-

cess conditions and using an Illinois No. 6 coal source is the U.S. Department of Energy Report DOE/MC/16372-3.

4.2.1 Moving Bed Gasification (Lurgi Dry Bottom)

The Lurgi Dry Bottom gasifier had been selected to typify a moving bed gasifier. The Lurgi pressure gasifier can probably be better described as a slow downward moving bed reactor where coal is successively preheated, dried, devolatilized, and the resultant char gasified with steam. The heat required is provided by oxidation of carbon with oxygen in a shallow combination zone near the bottom of the bed. The gases produced and released pass countercurrently upward through the bed.

The oxidant (98 mol percent oxygen) and the gasification steam are fed upward through a rotating grate upon which the downward moving ash collects before discharge through the gasifier bottom. The oxidant and steam are preheated as they pass through the grate and hot ash, which are thereby cooled. The flow of oxidant is regulated to achieve essentially complete gasification of the coal while the steam rate is controlled to limit temperatures in the fuel bed to a level below that at which the ash will fuse and clinker.

The Lurgi gasifier is classified as a low temperature gasifier with average gasification reaction temperatures in the range of 700 to 850°C. The heat balance for the Lurgi gasifiers of this study (cases ML and MH) are predicted upon a predicted hot crude gas temperature leaving the top of the bed of 620°C and an ash temperature leaving the gasifier of 480°C.

The predicted crude gas leaving the Lurgi gasifier proper at 620°C and 2240 kPa gauge has the following composition which forms the basis for material balances of this study:

Component	Mole % (wet)	Weight % (wet)
CH_4	4.28	
C_2H_4	0.13	
C_2H_6	0.20	
H_2	20.92	
CO	7.49	
CO_2	15.28	
H_2S	0.59	
COS	0.03	
N_2	0.18	
NH_3	0.40	
H_2O	50.50	
	100.00	98.06
Naphtha, tars and oils		1.64
Phenols and others		0.30
		100.00

The mixtures identified as naphtha, tars and oils and as phenols and others have the following analyses on a weight basis:

	Naphtha, Tars and Oils	Phenols & Others
Carbon	85.80	74.10
Hydrogen	6.90	6.40
Oxygen	4.25	17.00
Nitrogen	1.12	1.
Sulphur	1.93	1.50
	100.00	100.00

Further identification of the streams which are combined as naphtha, tars and oils, and as phenols and others is not available in the public domain. These materials have been characterized by the process licensor to the extent necessary for them to handle separate and recover them as part of the licensor's proprietary process package. The quantity of the combined recovery streams

and its heating value is essentially the only remaining characteristic available in the public domain.

4.2.2. Fluidized Bed Gasifier (Westinghouse)

The Westinghouse single state pressurized fluidized bed reactor has been selected to typify a fluidized gas gasifier. Pulverized (minus 6 mm size), dried (5 percent surface moisture) coal is pneumatically transported to the gasifier using recycled product gas. The coal, along with steam and oxygen is fed to the reactor through a central feed tube assembly. Combustion of part of the feed coal provides the heat needed to devolatilize the remaining coal to form a char and to react the char with steam producing hydrogen and carbon monoxide.

As the bed of char circulates through the jet formed at the end of the feed tube, the carbon in the char is combust d and gasified, leaving particles that are rich in ash. The fluid dynamic design of the gasifier allows the ash-rich particles which defluidize from the bed and collect in the ash annulus. The agglomerated particles are cooled by recycle product gas and removed from the gasifier through a lock hopper system. The raw product gas released from top of the fluidized bed passes through cyclones to recover entrained particles, largely char, for recycling to the gasifier.

The Westinghouse gasifier is classified as a moderate temperature gasifier with freeboard temperatures in range of 850 to 1050°C. The heat balance for the Westinghouse gasifiers of this study (cases FL and FH) are predicted upon a selected hot crude gas temperature over the fluidized bed of 1010°C and an ash temperature leaving the gasifier of 150°C.

The predicted crude gas leaving the Westinghouse gasifier proper at 1010°C and 2230 kPa gauge has the following composition which forms the basis for material balances of this study:

Component	Mole % (wet)
CH_4	6.53
H_2	26.14
CO	33.85
CO_2	13.09
H_2S plus COS	1.14
N_2	0.04
NH_3	0.10
H_2O	18.75
	100.00

4.2.3 Entrained Flow Gasifier (Texaco)

The Texaco Coal Gasification Process has been selected to typify an entrained flow gasifier. The overall Texaco gasifier essentially is a vertical refractory lined vessel comprising an upper reaction section where a coal-water mixture is partially combusted with oxygen, and a lower section where the bulk of the entrained slag is separated by flow reversal. The lower section contains radiant water-cooled surface to cool the crude gas. The coal is fed to the special designed burner, located in the top of the upper section, as a slurry of finely pulverized coal in water.

The upper or gasifier section operates at high pressure and high enough temperature for the ash to melt and form a free flowing molten slag. The coal slurry and oxygen combine at the burner. Part of the coal burns with the oxygen to produce a hot flue gas which provides the heat required for the endothermic steam/carbon and CO/CO_2 reactions.

Most of the ash, in the form of slag, falls into a water quench at the bottom of the radiant cooler portion of the gasifier and is withdrawn as a slurry through lockhoppers.

The Texaco gasifier is classified as a high temperature gasifier with gas temperatures leaving the gasifier portion in the range of 1250 to 1550°C. The heat balance for the Texaco gasifiers of this study (cases EL and EH) are predicated upon a predicted hot crude gas and an ash temperature of 1330°C leaving the gasifier zone.

The predicted crude gas leaving the Texaco gasifier proper at 1330°C and 4140 kPa, gauge, has the following composition which forms the basis for the material balances of this study:

Component	Mole % (wet)
CH_4	0.08
H_2	28.84
CO	42.45
CO_2	8.71
H_2S	1.01
COS	0.06
N_2	0.78
NH_3	0.19
H_2O	17.88
	100.00

4.3 Coal Derived Fuel Gas Specification

Historically, high caloric value natural gas has been the principal gaseous fuel used in combustion turbines. Its clean burning characteristics and generally ready availability made it ideal for the service.

Coal Derived Gas (DRG) fuels are basically good fuels and are for the most part acceptable. The particular fuel selected based on the specification and contaminate level will determine the life expectancy of the plant, as is the case with petroleum derived fuels. It appears from recent testing that component life for operation with either coal-derived or petroleum-derived fuels is comparable for fuels with like specifications. It can be stated that for

coal-derived gaseous fuel whose specification closely parallels the fuel presently used (i.e. natural gas) it is possible to design a high availability gas turbine combined cycle plant.

Contaminate levels in the coal gas fuels must be maintained to the point that particulates do not exceed six microns in size with the average particle size not exceeding three microns. As more knowledge becomes available on the corrosion-erosion and deposition effects of the coal derived fuels, the fuel will be selected to provide the best availability.

Tests (Reference 1) conducted by Westinghouse and the U.S. Energy Research and Development Administration at Lester, Pa., with full scale industrial gas turbine combustors have determined the following combustion characteristics:

Coal gas having a LHV of 130 Btu/sef (4845 KJ/m^3) burns stably, except in the operating region near idle. Coal gas of 200 Btu (7450 KJ/m^3), heating value burns acceptably throughout the engine operating range.

There is no NO_x problem with 130 Btu (4,845 KJ/m^3) gas. Emissions are probably acceptable with 200 Btu (7,450 KH/m^3), but will be unacceptable for 300 Btu (11,180 KJ/m^3) coal gas.

4.3.1 Heating Value

For the purposes of this study the fuel gases produced by the three modes of gasification with either cold or hot purification of the crude gas are used in mathematical model representative of a generic combustion turbine inclusive of its associated air compressor and combustor. Thus, differences due to the different gasification and purification modes are not obscured by attempting to match the combustion turbine requirements of each case to specific commercially available turbines. For each case it is presumed that an air

compressor-combustor-turbine combination could be designed which would have the performance characteristics on which the mathematical model was based.

Consequently, no attempt is made in this study to produce for any of the cases a fuel gas having a particular heating value, a particular specific gravity, gas or even a particular "gas index (GI)". The gas index is defined by the ratio of the heating value of the fuel to the square root of the specific gravity. This ratio, sometimes called the Wobbe number, is an index of the thermal input to a burner at a given pressure and therefore, is one indication of the interchangeability of different fuel gases.

The calculated lower and higher heating values at 25°C, the specific gravities relative to air, and the gas indices for the fuel gases of this study are given in Table 4-2. All are on a dry basis.

4.3.2 Particulate Matter

Particulate matter whether liquid or solid, if present excessively, can be damaging to the combustion gas turbine as well as being detrimental to the atmosphere.

To eliminate liquid particles the fuel gas in each of the low temperature purification cases is heated to at least 10°C above its saturation temperature.

For this study, the solid materials in the fuel gas to the combustion turbines has been limited to 30 ppm by weight. This is the maximum level recommended by General Electric. When low temperature purification of the fuel gas (cases ML, FL and EL) is practiced, it is expected that the quenching and subsequent scrubbing to remove sulphur compounds achieves removal to this level or less. It is also assumed for the high temperature purification cases (MH, FH, and

EL) that the mechanical removal means proposed for this study will achieve the desired reduction.

Such a dust loading (30 ppm by weight) corresponds to a high of about 3.7 grams per million (10^6) kilojoules of coal for case MH to a low of about 1.7 for cases ML and FL. Expressed in other units the 30 ppm by weight corresponds to about 27 milligrams per normal cubic meter of fuel gas for all cases. In English units one pound per million Btu is equivalent to about 430 grams per million kilojoules and one grain per thousand standard cubic feet is equivalent to about 2.42 milligrams per normal cubic meter.

Annually, the particulate matter discharged to the atmosphere would range from about 22 tonnes for case FL to about 45 for case MH assuming 70 percent operating factor.

4.3.3 Composition

Specific limits on sulphur content of gas turbine fuel gas are usually imposed by gas turbine manufacturers to prevent corrosion of low temperature equipment for heat recovery from the exhaust gases and to avoid formation, if alkali metal contaminants are also present, of corrosive alkali, sulphates during combustion. It is also more expedient to limit suphur input in the fuel by using a removal process before the gas turbine. This is due to the combustion turbine exhaust containing about 3 times the volumetric flow as a steam plant for comparable heat input.

This study is based on the recovery as product sulphur of at least 90 percent of the sulphur contained in the coal processed. Consequently, assuming no sulphur associated with ash, slag, or liquid plant effluents, the total sulphur discharged to the atmosphere amounts to 10 percent or less of that originally present in the coal. This discharge to the atmosphere is primarily as flue gas from the combustor turbine heat recovery train though some

would be as off-gas from the sulphur train though some would be as off-gas from the sulphur recovery facilities.

For all cases, therefore, the SO_2 discharge to the atmosphere amounts to more than 0.26 kilograms per million kilojoules of coal gasified. This discharge rate would result in an annual discharge ranging from about 3150 tonnes of SO_2 for case EH to 3860 tonnes for case ML assuming a 70 percent operating factor.

4.3.4 Other Contaminants

A major concern relative to reliable turbine operation is the control of the contaminants that cause corrosion of hot section components. The alkali metals (Na and K) in addition to vanadium (Va) and lead are the principal components in fuels that cause hot corrosion.

The alloys, of which the nozzles and blades are made, are protected from high temperature oxidation by a dense adherent oxide film which excludes the excess oxygen in the combustion gases from gaining access to the metal itself. If this oxide film is removed, ruptured or modified such that the metal is exposed to the atmosphere, oxidation can proceed rapidly.

The above contaminants, when in sufficient concentration in the combustion products to deposit on the surface of the nozzles and blades, tend to disrupt these oxide films. This is accomplished both by dissolving the film and by reaction with the base metal. The alkali metals almost always exist in the form of sulphates: Na_2SO_4 and K_2SO_4. It is postulated that as these deposit, sulphur from them diffuses into the base metal alloy forming chromium sulphide thereby depleting chromium in the alloy. The net effect is that the protective film cannot re-form due to depletion of chromium from the surface layer. Instead a thick but porous scale forms which does not protect the base alloy from oxygen.

The oxide of vanadium V_2O_5, simply dissolves the oxide film permitting easy access of oxygen to the alloy. The oxides and sulphates of lead have a similar effect. The occurence of lead and vanadium in coal is so rare that they would not usually be of concern.

Coal contains both potassium and sodium alkali contaminants. They have been traditionally assumed to be equivalent in their aggressiveness and that to achieve a given corrosion life the total of Na plus K was limiting. Work by General Electric indicates, however, that potassium in combination with sodium is more aggressive than what would be predicted for the simple sum of Na and K.

General Electric defines an equivalent Na level, Xe, as that level of sodium which would give the same corrosion as a given (and smaller) combined level of Na plus K:

$$Xe = R\ (Na_e + K_e)$$

where R is a function of Na/K ratio in the exhaust gases, Na_e is the concentration of Na in the exhaust gas expressed in ppm by weight, and K_e is the concentration of K in exhaust gas.

The presence of alkali contaminants in sources other than the fuel, such as the air, must also be considered in predicting the allowable level in the fuel. With respect to sodium

$$Na_e = \frac{(A/F)\ Na_A + Na_F}{(A/F) + 1}$$

where

A/F is the air-to-fuel mass ratio used for the turbine
Na_F is the sodium concentration in the fuel
Na_A is the sodium concentration in the air
Na_e is the sodium concentration limit in the combustion products

A similar expression can be written for potassium.

In order to achieve a first stage blade life equivalent to that expected when a petroleum distillate fuel (with about 1 ppm sodium) is used General Electric indicates that the equivalent sodium concentration, X_1, should be about 0.02 ppm in the exhaust gas.

When low temperature purification of the fuel gas (cases ML, FL, and EL) is practiced the cooling, quenching and scrubbing to remove sulphur compounds effectively reduces the alkali content to levels well below that recommended.

On the basis of the predicted alkali concentration reported in Table 4-1 only one high temperature case, case MH, would be able to meet the General Electric recommendation without question. However, it must be recognized that very little data is available on the alkali metal content that can be realistically expected in the gas from commercially available gasification technologies.

Lacking such data it is impossible to verify the predictions made by a mathematical model. For all the other high temperature purification cases (FH, and EH) it has been assumed in this study that the alkali content of the flue gases will not cause excessive hot corrosion of the generic combustion turbine and will not require means to specifically remove the alkali content.

There are several factors which might be said to support or justify such an assumption. They include:

- the paucity of data to evaluate the accuracy of the predicted alkali content

- the model predicts that, overwhelmingly, the bulk of the volatile potassium and sodium would exist as the chlorides and the hydroxides. The predicted levels of sulphates for all cases are extremely low

- coatings, such as platinum - chromium - aluminide, can be used to extend the life of turbine components in moderately corrosive environments, or can permit the use of somewhat more contaminated fuels

- a shorter operating life, or in other words, higher maintenance costs may prove economically justifiable with high temperature purification

- indications by some investigators that particulate removal devices incorporating beds of granular or pelletized solids for removal of particulates might also serve to remove the alkalies by reacting with them. Some materials under consideration include kailin, bauxite and even ash from high temperature gasifiers.

While some of the above items would increase the capital and/or operating costs if they were necessary to make high temperature purification possible, no attempt has been made in this study to place a value on them. The absolute need for any of them, and the extent of that need would make incorporating them problematical.

TABLE 4-2

FUEL GAS CHARACTERISTICS (Dry Basis)

Case	ML	MH	FL	FH	EL	EH
Type Gasifier	Moving Bed		Fluidized Bed		Entrained Flow	
Purification Temperature Level	Low	High	Low	High	Low	High
Lower heating value @ 25°C, kJ/m³*	10,830	11,380	12,200	11,800	10,720	10,520
Higher heating value @ 25°C, kJ/m³*	12,610	12,650	13,190	12,750	11,430	11,230
Specific gravity relative to air	0.666	0.739	0.705	0.731	0.677	0.706
Gas index $(LHV \div (S.G.)^{\frac{1}{2}})$	13,260	13,230	14,530	13,800	13,030	12,520

*m^3 measured at normal conditions or 0°C 101.3 kPa absolute.

4.4 Combined Cycle System for Power Production

4.4.1 Capacity

Improvement in heat rates achievable by high temperature purification is not in itself sufficient to motivate development of means to commercially implement high temperature purification and particulate removal. It is also necessary to estimate the capital cost requirements of high versus low temperature purification because there will be no incentive to pursue the lower heat rates if the capital costs to the achieve them result in more expensive electric power. On the other hand, little or no improvement in heat rate might, if achievable at sufficiently lower capital cost, be economically attractive.

A plant capacity of 215 megawatts (net) was established for this Study as the basis for making the capital and operating cost estimates for the high and low temperature purification alternates for each gasification technology.

4.4.2 Combustion Turbine

A mathematical model representative of a generic combustion turbine, including its associated air compressor and combustor, is used to calculate the overall performance of the turbine-compressor system of each case. The use of a simplified thermodynamic computer simulation may not be precise but it is realistic and consistent with the requirements of a comparative study of hot versus old purification. The inaccuracies involved in such a simulation cancel out when calculating performance differences between cases employing different purification methods. Consequently, performance and efficiency differences thus calculated are attributable to purification system differences.

The thermodynamic design of the turbine-compressor system for each of the six cases is different primarily because the available fuel gas is different in each case. Other parameters such as the following were used for all cases:

- ambient conditions of 101.35 kPa absolute and 16°C
- turbine inlet temperature of 1200°C
- air compression ratio at ISO conditions (15°C) of 14
- isentropic efficiency of 88 and 90 percent, respectively, for air compressor and gas expander

- inlet air pressure drop of 15 millibars is allowed

- to allow for exhaust gas duct and heat recovery equipment losses 40 mbars is allowed for all cases except case ML where an additional 5 mbars is used to accommodate after-burning of recovered hydrocarbon liquids

- fuel gas is delivered to the combustion turbine trip valve at a pressure of 1900 kPa gauge

- approximately 4.5 percent of the compressor discharge air is cooled by heat exchange, to 175°C for cooling the turbine rotating hot parts. Heat removed by this exchange is recovered and integrated into the medium pressure section of the heat recovery facilities associated with the turbine

- the calculated generator outputs include a deduction for the mechanical and electrial losses of compressor-turbine-generator system. This deduction which is inclusive of lube and seal pumps, equals about 3.5 percent of the gross power output of the expander.

4.4.3 Heat Recovery Steam Generation

The heat recovery steam generation trains (HRSG) are conceptually the same for all cases except case ML in that they are structured to effectively recover heat in the form of steam only from the gas turbine exhaust gases. In case ML (moving bed gasification with low temperature purification) hydrocarbon liquids are recovered which are not suitable for direct firing of the gas turbine. They are conceptually fired by means of suitable burners into the turbine exhaust cases ahead of the HRSG.

The makeup of the various heat recovery trains depends upon the process relationships existing between the different gasification-purification combina-

tions and the combined cycle facilites, and upon the degree to which they can be effectively integrated. In all cases, an approach temperature of about 17°C (temperature of gas leaving minus saturation temperature of steam) has been used in calculating the steam generated in the evaporator sections. This approach at the pinch points is considered to provide a reasonable balance between cost of surface and steam production. Care has been taken to avoid the possibility of condensation occurring on the cooler surfaces in the HRSG.

4.4.4 Steam Turbine

Steam is generated at several pressure levels by heat recovery from the process units and from the exhaust gases of the combustion turbines. Conceptually, the process steam and the thermal energy needs of the plant are satisfied by suitable letdown turbines driving the compressors of the air separation units and the boiler feedwater pumps. Ultimately, the steam, not consumed by the process and for heating, is used in an appropriate condensing turbine to drive an electric generator. The electric power generated is used for the other mechanical drivers of the plant, with the remainder together with that produced by the combustion turbine driven generators, exported.

For purposes of this Study, the selection of specific commercial steam turbines has not been attempted. Instead the power developed by expansion or letdown of steam in the turbines has been calculated in all cases by applying an overall efficiency factor of 0.8 to the appropriate isentropic or theoretical steam rate. In all cases, the steam condenser condition used is 63.5 mm of mercury absolute. The calculated electric generator outputs include a deduction of 2.0 percent for the mechanical and electrical losses of the turbine-generator system.

While this simplified simulation may not be precise it is realistic and is consistent with requirements of a comparative study of hot versus cold purification. The inaccuracies involved largely cancel out in calculating perfor-

mance differences between cases employing different purification methods. Consequently, performance differences thus calculated are attributable to purification system differences.

5.0 Review of Low Temperature Clean-up Technology

5.1 Gasification System and 'Clean-up' Technology

5.1.1 Moving Bed-Lurgi (Low Gasification Temperature)

5.1.1.1 System Operation

In a moving bed, high pressure 2.3-3 Mpa (350-450 psig) gasifier, coal closely sized to obtain 3 mm x 37.5 mm (1/8" x 1 1/2"), enters the gasifier back hopper. Coal fines generated during crushing and screening are removed from the system. The coal is fed into the top of the reactor and flows in a counterwise (downward) direction to the gas, steam and air (or oxygen) that are introduced at the bottom of the gasifier to cause coal gasification reactions. As the steam and oxygen pass upward four different zones can be characterized in the coal bed by the prevailing reactions and temperatures. They are, from bottom to top, carbon combustion, gasification, devolatilization and drying. A high temperature zone 955-1149°C (1750-2100°F) occurs in the combustion zone and becomes progressively cooler in the endothermic gasification and devolatilization zones. Depending on the coal type and operating conditions the temperature of the outlet gas will range between 300°C (570°F) for lignite with high moisture content and 650°C (1,200°F) for coals with a low reactivity and low moisture content. A water cooled, rotating grate is sometimes used to prevent clinkering and to facilitate ash removal.

The crude gas formed in the gasifier contains carbonization products such as tar, oil, naphtha, phenols, ammonia, etc. and traces of coal and ash dust. This crude gas is passed through a scrubber where it is washed by circulating gas liquor and then cooled to a temperature at which the gas is saturated with steam. As higher boiling tar fractions are condensed, the wash water contains tar to which coal and ash dust is bonded.

5.0 Review of Low Temperature Clean-up Technology (Cont'd)

5.1.1.1 System Operation (Cont'd)

The stream-saturated gas is passed to a waste heat boiler in which waste heat at a temperature of 160°C (320°F) to 182°C (360°F) is recovered. The gas liquor condensed in this boiler is pumped to the scrubber and sulphur gas liquor is routed to a tar-gas-liquor separator. The mixture of tar and dust is returned to the gasifier for cracking and gasification.

Approximately 86 percent of the coal fed to the gasifier is gasified and the balance of 14 percent which is mostly carbon is burned in the combustion zone with oxygen. Thus, the latent heat of this carbon is converted to sensible heat, which is transferred to the gasification agent flowing upstream counter-currently to the coal. This technique makes it possible to perform complete gasification of the coal.

5.1.1.2 Gas Clean-Up System

The product of gasification is a mixture of various components, mainly CO_2, CO, CH_4, H_2, H_2O, and products of coal carbonization, such as tar, oil, naphtha, phenols, etc., plus sulphur compounds and other impurities. The particulate and tar in the gas has been removed in the scrubber. The composition of the gas is determined by the following:

a) The pressure and temperature of the gasification process and the composition of the gasification agent, which consists of 90 percent by volume of oxygen as a source of heat supply for the endothermic gasification.

5.1.1.2 Gas Clean-Up System

b) The type of coal

1) The reactivity of the coal determines the consumption of oxygen and the degree of approach to the gas equilibrium.

2) The ash behaviour basically determines the steam/oxygen ratio of the gasification agent.

3) The volatile matter influences the composition of the gas impurities, and these such as sulphur, nitrogen and chlorine determine the amount of impurities in the gas.

The gas composition of the gas leaving the gasifier is determined by the optimum operating conditions of the gasifier. The resulting H_2/CO ratio of the gas must be altered for the methane synthesis required for synthetic natural gas (SNG). Therefore, a shift conversion reaction ($CO_2 + H_2O = CO_2 + H_2$) is necessary to adjust the H_2/CO ratio. The undecomposed steam in the crude gas downstream of the waste heat boiler is used in the gasification section for the performance of the shift reaction.

Purification of gas from coal gasification to the degree of purity required for SNG is difficult due to the large and varied quantity of impurities. The Rectinsol process was developed for this application, and is described in detail in Section 5.2.2.1. The Rectinsol unit for an SNG plant has the following three sections:

1) Prewash for the removal of gas naphtha, unsaturated hydrocarbons and other high boiling gas impurities.

5.1.1.2 Gas Clean-Up System (Cont'd)

2) Removal of CO_2, H_2S and COS. The gas leaving this section is the purity required for the methane synthesis.

3) Removal of last portion of CO_2 and dehydration of the gas.

All the impurities contained in the gas can be completely removed by the Rectinsol process. This can be demonstrated in many large commercial installations.

The Rectinsol installations have demonstrated that a purity of less than 0.1 ppm total sulphur can be continuously maintained at stream factors of more than 95%.

In the above gasification process, a gas liquor is produced from the fairly large amount of steam which is condensed in the various processing steps. The gas liquor has been in contact with coal carbonization products and contains soluble compounds mainly phenols, ammonia and fatty acids. The Lurgi Phenosolvan Process removes the phenols and the ammonia, leaving an effluent with amounts below 20 ppm phenols and below 60 ppm free ammonia. This can be further processed to meet regulations for the disposal of waste water.

The conclusion drawn from the above is that the Lurgi gasifier with the Rectinsol and Phenosolvan processes produces a synthetic natural gas (SNG) that can meet the Gas Turbine Manufacturers Standards for coal gas fuels and also meet environmental standards.

5.1.1.3 Influence of Coal Properties on the Gas Clean-Up System

Ash

Commercial experience has shown that ash is one of the main concerns. It is almost impossible to predict the gasification behaviour of any ash from a standard analytical examination. In the Lurgi gasifier the ash and various ash components are distributed within fuel bed and are subjected to various decomposition reactions in a changing atmosphere. It therefore requires a coal test in a gasifier to define actual ash behaviour.

a) Lignite Type Ash

North Dakota Lignite - analytical data

$$\frac{CaO + MgO}{Al_2O_3 + SiO_2} = 0.74 \ (-\ 1.19)$$

$Na_2O + K_2O = 6.5 \ (-\ 14)$ % b.w.

	Reducing atm.	Oxydizing atm.
Softening point	1190°C (2190°F)	1238°C (2260°F)
Melting point	1209°C (2208°F)	1238°C (2260°F)
Fluid point	1210°C (2210°F)	1244°C (2272°F)

The above data are indicative of "short slag" and suggest that there may be ash problems.

5.1.1.3 Influence of Coal Properties on Gas Clean-Up System

The lignite was successfully gasified in a large scale test at Sasolburg, South Africa. The high reactivity of the lignite allows the Lurgi gasifier to be operated at a steam-oxygen ratio (low temperature, approximately 649°C [1200°F]. The operation was stable and very successful and furnished an ash which was entirely different from what the analyses had indicated. It consisted of 85% fines which were slightly sintered, the balance being clinker which was easily broken by the grate. The amount of dust was relatively low. These ash characteristics are a good basis for trouble-free operation.

b) Acid Ashes with High Pyrite Content

The Eastern U.S. coals yield ash with quite different properties. The following is a typical analysis of Illinois coal processed in Westfield, U.K.:

$$\frac{CaO + MgO}{Al_2O_3 + SiO_2} = 0.105$$

Fe_2O_3 = 22.0% b.w.

	Reducing atm.	Oxydizing atm.
Softening point	1010°C (1940°F)	1310°C (2390°F)
Melting point	1071°C (1960°F)	1410°C (2570°F)
Fluid point	1131°C (2067°F)	1418°C (2585°F)

The high pyrite content explains the variation in fusion properties for oxidizing and reducing atmospheres.

The Westfield tests have showed that coal with this type of ash is well suitable for gasification.

It is apparent from the above that the Lurgi gasifier is capable of gasifying a range of lignite and coal and that the amount of dust released is very low.

5.0 Review of Low Temperature Clean-up Technology (Cont'd)

5.1 Gasification System and Clean-Up Technology (Cont'd)

5.1.2 Fluidized Bed - Westinghouse (Medium Gasification Temperature)

5.1.2.1 System Operation

The Westinghouse Electric Corporation in the last decade has developed a pressurized, fluidized bed gasification process for the production of low and intermediate Btu fuel gas. The process development unit (PDU) since 1975 has had more than 8,000 hours hot operation. A wide variety of coals, lignites and chars. have been tested at capacities ranging from 13,586 Kg per day (15 tpd) with air to 31,703 Kg per day (35 tpd) with oxygen.

Run-of-mine coal is received in lump form and prepared in Williams impact mill crusher-dryer, providing a feedstock size of 6 mm + 0 (-1/4 inch + 0) with a surface moisture content betwen 5 and 10 percent.

Coal feeding is performed by pressurized lockhoppers and pneumatic transport of dry feedstock using recycled product gas as the transport medium. The feed material is batched from storage bins by belt conveyors and bucket elevators to a transfer lockhopper that cycles between atmospheric pressure and system pressure. Batches of coal are then transferred to the pressurized feed lockhopper, where a rotary feed valve discharges the material into a pneumatic transport feed line.

The gasifier, which is an ash agglomerating fluidized bed, can process run-of-mine coal with steam and oxidant (air and oxygen) to produce a low-to-medium-Blu gas and ash agglomerates. Typical operating ranges for this unit are as follows: freeboard temperature - 899 to 1010°C (1650 to 1850°F), pressure 0.90 Mpa to 1.59 Mpa (130 to

5.0 Review of Low Temperature Clean-up Technology (Cont'd)

5.1.2 Fluidized Bed - Westinghouse (Medium Gasification Temperature) (Cont'd)

5.1.2.1 System Operation (Cont'd)

230 psig, coal feed rate - 453 to 725 Kg/hr (1000 to 1600 lb/hr) for bituminous coals, 725 to 1585 Kg/hr (1600 to 3500 lb/hr) for sub-bituminous coal and for lignites.

The reactor vessel consists of four functional zones: (1) an expanded freeboard section to separate char fines from exiting raw gas; (2) a gasifier bed to permit char particles to react with steam; (3) a combustor to devolatilize coal, combust carbon for the heat source, and produce ash agglomerates; and (4) a char-ash separator to separate ash agglomerates from char, to cool the ash with recycle gas and to discharge solids into the lockhoppers.

Coal and recycled fines are injected by an axially placed feed tube in a pneumatic conveying mode using recycle gas. Oxygen and steam are preheated to 260 °C (500°F) and are also introduced in the feed tube. A lean-phase turbulent jet forms at the exit of the feed tube assembly. Coal is rapidly devolatilized in an oxidizing environment; the jet induces a vigorous recirculation of solids (char) and promotes heat transfer into the upper regions of the fluidized bed.

The ash produced in the gasifier is removed in a dry-granular form (agglomerates). The discharge is controlled by a starwheel feeder, which feeds a set of parallel pressurized lockhoppers, mounted at the bottom of the fluidized bed.

5.1.2.2 Gas Clean-up System

Fines-laden raw gas from the gasifier flows through two cyclones, which separate between 80 to 90 percent of the elutriated char, and then to a quench scrubber, where the gas is cooled to its adiabatic saturation temperature by water spray. Fines from the primary cyclone are cooled by recycle gas and pneumatically transported to a desengaging lockhopper. The primary fines are then pressure boosted using a two-lockhopper scheme and recycled into the gasifier. The secondary fines fall through a dipleg and are cooled and recycled directly without lockhoppers.

A slipstream of quenched gas is further scrubbed in a packed bed contact cooler and recompressed for recycle to the gasifier. This recycle gas is used for pneumatic transport of solids, ash cooling and separation and gasifier grid aeration. The bulk of the product gas is incinerated in a thermal oxidizer.

The water-fines slurry from the scrubber bottoms is processed through a clarifier pit and the clear water is recirculated by high pressure pumps into the scrubbers. Sludge from the pit is periodically removed to a contract landfill facility.

Ancillary units include an oxygen storage dewar, air compressor and an electic steam generator. All systems are extensively instrumented to monitor and record process performance data. Gas characteristics equipment includes two on-line gas chromatographs for the analysis of product gas trace and bulk constituents, an infrared analyzer, a calorimeter, and a gas chromatograph specially devised for estimating the moisture content in the raw gas exiting the gasifier. Fines entrained by the gas are measured by isokinetic probes and sample conditioning trains. Digital data are recorded by a computer at a scan rate of 10 seconds.

5.0 Review of Low Temperature Clean-up Technology (Cont'd)

5.1.3 Entrained Bed - Texaco (High Gasification Temperature)

5.1.3.1 System Operation

The following is based on the Ruhrkohle/Ruhrchemie (pilot) demonstration coal gasification plant located in Oberhausen-Holten, West Germany. This plant contains a Texaco gasifier and has a nominal coal capacity of 5,700 to 6,100 Kg/hr (150-160 tpd).

In the plant, coal is fed from a storage bunker to a grinding mill. There it is pulverized and combined with fresh water or, alternatively, effluent water recycled from the settler. The resulting coal slurry is pumped to agitated run tanks and from there into the gasification reactor.

The Texaco Coal Gasification Process reactor is a pressurized, entrained bed, downflow slagging gasifier. It operates under pressures of 2 to 10 Mpa (300 to 1,500 psia) and at high temperatures, generally in the range of 1,200 to 1,400°C. These temperatures are above the melting point of the coal ash.

The coal slurry and oxygen are fed into the top of the gasification reactor vessel. This vessel is lined with refractory, and it consists of two zones, a gasification zone and a quenching zone. The basic combustion and gasification reactions occur in the gasification section. The raw synthesis gas from the gasification section passes into the radiant cooler or quench section of the vessel. Here the gas is partially cooled and process steam generated.

To date the Westinghouse process and operating data base extends to more than 8,000 hours of hot operation with coals and chars in the

5.1.3.1 System Operation (Cont'd)

air-blown mode to produce low-Btu gas and using oxygen to produce medium-Btu gas. Table 1 lists the feedstocks that have been processed in the unit.

TABLE 1

FEEDSTOCKS PROCESSED IN WESTINGHOUSE GASIFIER

Lignite:	Texas Lignite
Subbituminous Coals:	Wyoming Wyodak, Montana Rosebud, Republic of South Africa Coal
Bituminous Coals:	Pittsburgh No. 8, Indiana No. 7, West Virginia, Illinois No. 6, Ohio No. 9
Chars:	Metallurgical Coke Breeze, Petroleum Cokes FMC (COED) Char, Devolatilizer Chars

5.1.3.2 Gas Clean-Up System

Synthesis gas from the gasifier is routed into a quench tower, where recycled process water is used to additionally cool the gas and remove particulate matter. The synthesis gas leaves the quench tower and passes into a water scrubber where the final gas cooling and particulate removal takes place.

5.1.3.2 Gas Clean-Up System (Cont'd)

Most of the slag produced in the reactor is solidified in a water bath in the bottom of the radiant cooler section. It is then removed from the reactor through an electronically-controlled lockhopper system. The slag and sluice water fall into a covered bin where the granulated slag is continuously removed by a conveyor belt. The slag water, containing some fine particulates, is sent to the settler. The water from the radiant cooler section of the gasifier and from the quench tower is routed to a flash tank where dissolved gases are flashed off during depressurization. The water from the flash tank then passes into the settler. The residence time in the settler is sufficient to alow settling of most of the fine particulate matter.

Two water streams are taken from the settler. The overflow water stream, taken off near the top of the settler, contains a relatively low concentration of solids. The underflow stream, taken from the bottom of the settler, contains the residual fines at a much higher concentration.

The settler overflow is combined with the scrubber blowdown stream and makeup water stream. This combined water stream is used as slag sluice water and also recycled to the quench tower and the quench zone of the gasification reactor.

The underflow (or at least a portion of it) from the settler serves as the water blowdown stream from the coal gasification unit. The blowdown stream is first sent to an open holding tank before being routed to the plant water treatment system.

5.2 Sulphur Removal Technology

5.2.1 Factors Considered in Technology Selection

The primary objective of the Sulphur Removal System is the purification of the fuel gas by removing undesirable components which are primarily H_2O and COS. The composition of the fuel gas being purified is dependent on the characteristics of the coal being gasified, the gasification system, and its operating conditions.

5.2.1.1 Ability to Remove Sulphur Compounds

The single most important factor in choosing a Sulphur Removal System is its ability to remove sulphur compounds effectively and efficiently from the fuel gas sufficiently to satisfy its end use requirements.

5.2.1.2 Selectivity

The selectivity of the solvent used in the Sulphur Removal System for sulphur compounds over other components in the crude fuel gas is an important factor. Ideally the solvent would effectively remove the sulphur compounds to the required extent without simultaneously removing other components. While loss of combustible components is obviously very undesirable, the loss of other components, such as CO_2, is also undesirable because it results in reducing the mass flow through the gas turbine. Minimizing the loss of C_2O is difficult because CO_2 and H_2S are similar in their properties both being classified as acid gases.

A more selective solvent results in a higher sulphur compound content in the off-gas produced when regenerating the solvent. This inherently provides greater flexibility in the selection of the

5.2.1.2 Selectivity (Cont'd)

sulphur recovery process and makes more suitable the use of the conventional cost-effective Claus recovery system.

5.2.1.3 Energy Requirements

An important factor that must be considered in the choice of a Sulphur Removal System is its energy requirements. Many of the processes available use thermal energy, usually as steam, to regenerate the absorbing solvent, and the quantity of heat required can be significant. Additionally, acid-gas removal is usually effected under high pressure conditions and solution regeneration is conducted at low pressure. This will necessitate regenerated lean solvent pumping to the absorption tower. The power requirements can be appreciable depending on solvent circulation. The degree of process selectivity for sulphur compounds and amount of sulphur removal will impact greatly on the system's utilities. Finally, refrigeration of the solvent, if needed, would add to the capital cost and the utility requirements.

5.2.1.4 Contaminants

Coal is a very complex feedstock and the fuel gases produced from it can potentially contain a wide variety of chemical compounds and elements as trace components. Their nature and the quantities of them would, of course, be dependent not only on the coal, but also on the method and conditions used for its gasification. Prediction of the presence and identify of such contaminants is difficult as is their actual detection. Only actual experience of the sulphur removal system in association with coal gasification process will

5.2.1.4 Contaminants (Cont'd)

prove the true compatibility of the processes. The cyclic nature of the sulphur removal systems means that essentially undetectable levels of a contaminant in the gas may with time accumulate to significant levels in the solvent.

The use of methanol scrubbing, or Rectisol, for sulphur removal at the AECI coal-based ammonia plant near Johannesburg, South Africa, experienced considerable operating difficulties with the gas from Koppers-Totzek entrained gasifiers. Such was unexpected and has been attributed to the presence of trace amounts of oxygen, nitrogen oxides, and sulphur dioxide in the gas being scrubbed. Low temperature catalytic hydrogenation to remove these impurities has, it is reported, eliminated the problem.

5.2.1.5 Environmental Considerations

The direct discharge to the environment of effluents from the Sulphur Removal Systems is normally non-existent. Under some circumstances a liquid effluent stream is required to maintain a water balance for the system. Such a stream normally is processed as part of the Sulphur Removal System so as to minimize the presence and, therefore, the loss of the solvent.

Almost all Sulphur Removal Systems incorporate some degree of storage facilities for the solvent. These facilities used a source of makeup and may be used for solvent retention during periods of maintenance. Solvents of low volatility are preferred because the breathing and other losses of solvent vapors is less and the chance of detrimental environmental impact is reduced.

5.2.1.5 Environmental Considerations (Cont'd)

The possibility of massive leaks and other operating upsets may involve at times the discharge of solvent as a liquid or vapor under sometimes uncontrolled conditions. These possibilities make use of non-toxic and non-hazardous solvents preferable. The considerable inventory of solvent in a Sulphur Removal System makes use of a non-combustible one preferred if possible.

5.2.1.6 Capital Costs

The capital cost for any sulphur removal technology will be a function of the gasifier with which it is associated. The sulphur removal technology requiring the lowest capital cost may be different for different gasification technologies. Therefore, while contributing to the cost of the electric power being generated, the capital cost of the sulphur removal facilities by itself will represent only about 10 percent of the overall cost of the combined cycle facilities no matter which process is selected. Of greater importance than the capital cost is the capability of the selected system to fulfill the process specifications required of it without the use of excessive utilities; without undue adverse effect on capital and operating costs of the sulphur recovery facilities, without adversely impacting on the heat rate of the plant through excessive loss of combustibles and carbon dioxide, and without adverse impact on the environment.

5.2.1.7 Other Factors for Consideration

Many other factors can influence the choice of a Sulphur Removal System including:

5.2.1.7 Other Factors for Consideration (Cont'd)

- Corrosiveness of solvent
- Availability of the solvent and its cost
- Experience with the solvent
- Physical characteristics such as freezing point, viscosity, specific gravity, and foaming tendencies

5.2.2 Sulphur Removal Technology Review

The Table 5-1 presents a listing of many of the systems that might be employed to purify fuel gas produced from the gasification of coal. The systems are classified by the general type of process and the table represents a qualitative assessment of the relative selectivity and energy requirements of the various systems. It is, by no means, an exhaustive listing of all available processes but does present a fair representation of the currently available systems which are usually considered.

Amine solvents are generally less selective than physical solvents and have higher energy requirements for regeneration. MEA and DEA suffer from both degradation by COS and O_2, and vaporization losses. DGA and DIPA are less volatile amines but are only marginally better for acid gas removal at low system pressures. DIPA has partial selectivity and is a likely candidate for concentrating acid gases at low pressures prior to sulphur recovery. Alternative systems using other than amine solvents are generally employed at greater than one atmosphere total partial pressure of acid gas (CO_2 plus H_2S).

The modern hot carbonate processes are basically an outgrowth of low pressure carbonate systems that were developed in the 1920's and 1930's for acid gas removal. Acid gases are absorbed in a counter-

TABLE 5-1
ACID-GAS SYSTEMS EMPLOYED IN COAL GASIFICATION[a]

Solvents	General operating range	Selectivity[a]		Component distribution				Utilities[b]
	CO_2 partial press.	H_2S/CO_2	CO_2/HC	NH_3	COS	RSH	Other organics	Range dependent on conditions (partial pressure CO_2 and degree of cleanup)
Amines								
MEA	Low	1	8	c,f	g	c,d,f	c,f	10
DEA	Low	1	8	c,f	c,d	c,d,f	c,f	10
DGA (Fluor)	Low	1	8	c,f	c,d	c,d,f	c,f	8
MDEA	Low	3	8	c,f	c,d	c,d,f	c,f	7
TEA	Low	3	8	c,f	c,d	c,d,f	c,f	7
DIPA (Shell)	Low	2	8	c,f	c,d	c,d,f	c,f	7
Hot Carbonate								
Catacarb (Eickmeyer and Associates)	Moderate	2	9	c,f	h	f,i	c,f	4
Benfield (Benfield Corp)	Moderate	6	9	c,f	h	f,i	c,f	4
Alkazid (I.G. Farbin Industrie, W. Germany)	Moderate	3	9	c,f	h	f,i	c,f	4
Giammarco-Vetrocoke	Moderate	9	9	c,f	h	f,i	c,f	4
Physical Solvents								
Methanol (Rectisol-Lurgi)	High	7	2	f,e	c,d	c,e,f	c,d,e,f	2
DMEPED (Selexol-Allied)	High	9	3	f,e	c,d	c,e,f	c,d,e,f	1
Propylene carbonate (Fluor)	High	4	3	f,c	c,d	c,f	c,d,e	1
N-methyl pyrrodlidone (Purisol-Lurgi GAF)	High	9	3	f,e	c,d	c,e,f	c,d,e,f	1
Tri-n-butylphosphate (Estasolvan)	High	4	3	f,c	c,d	c,f	c,d,f	1
Methyl cyanoacetate (Union Oil)	High	4	5	f,c	c,d	c,f	c,d,f	1
Mixed Solvents								
Sulfolane + DIPA (Sulfinol, Shell)	Moderate	2	3	f,c	c,d	c,f	c,d,f	5
Methanol + DGA (Amisol, Lurgi)	Moderate	1	3	f,c	c,d	c,f	c,d,f	5

[a]Arbitrary number scale 1-10; 1 = poor selectivity, 10 = excellent selectivity
[b]Arbitrary number scale 1-10; 1 = low utilities, 10 = high utilities
[c]With CO_2-H_2S stream, after simultaneous removal
[d]With CO_2 stream from bulk removal following selective H_2S removal
[e]With H_2S stream from selective H_2S removal
[f]Separate stream prior to CO_2-H_2S removal
[g]Degrades solvent
[h]Hydrolyzes to H_2S
[i]Largely not removed; remains in treated gas
[j]Institut Francais du Petrole: Friederich Uhde, G.m.b.H.

[a]Fleming, Donald K., "Acid-Gas Removal Systems in Coal Gasification", Tennessee Valley Authority Ammonia from Coal Symposium, May 8-10, 1979, (Table presented in paper)

current contactor by a hot carbonate solution under moderate pressure conditions. The gases are recovered from the rich solution by flashing and steam stripping in a low pressure regenerator. The processes are used extensively for the removal of CO_2 and sulphur compounds in the natural gas, chemical, and refining industries.

The modern carbonate systems employ a 20%-30% water solution of potassium carbonate, operated between 220°F and 300°F, and utilize various additives to increase the rate of gas absorption as well as to catalyze the solution reactions. Advantages for these systems include lower regeneration energy requirements than for amine systems, lower hydrocarbon absorption losses than for organic physical solvents, the ability to handle carbon disulphide (CS_2) and carbonyl sulphide (COS) through hydroloysis to H_2S, and minimal degradation of the carbonate solution. A degree of H_2S selectivity can be achieved by proper design of the system. Negative features are the highly corrosive atmosphere provided by the solvent and the higher relative utilities required in comparison to the physical solvents. Some of the "hot pot" processes utilize additives, for example arsenic salts, which are toxic.

The third general method for removing acid gases from a process gas stream is by physical absorption in an organic solvent without chemical reaction. The solvent can be regenerated by heat, pressure reduction, or inert gas stripping, and can produce a concentrated stream of adsorbed acid gas. H_2S and CO_2, as well as minor components in the gas stream (including CS_2 and mercaptans) are more soluble in organic solvents than fuel gas species, especially at elevated pressure. Physical solvent systems' benefits include lower heat and power consumption than other systems discussed, a lower

corrosion rate and the ability to remove minor impurities from the fuel gas stream. More importantly, some physical solvents exhibit high selectivity for H_2S over CO_2. This feature will favorably impact the sulphur recovery system selection, the regeneration energy requirements and the final composition of the fuel gas. Major disadvantages of the physical solvents are possible hydrocarbons losses from the fuel gas for some processes, combined with high solvent cost and losses from the system.

Several commercially available sulphur removal or acid gas removal technologies which are considered representative of the general types of process discussed above, are described in further detail below. The sulphur removal systems selected for further elaboration are the Rectisol Process which both employ physical solvents, the Benfield Hipure Process utilizing a hot potassium solution as solvent, and the Fluor Econamine Process using the amine solvent diglycolamine.

5.2.2.1 Rectisol Process

The Rectisol Process is a proprietary process, licensed by both Lurgi GMBH and Linde A.G., based on the physical absorption of acid gases in cold methanol. The process is in commercial operation at 36 operating plants including several coal gasification facilities. The system draws its principal advantage from the different solubilities of various gases in methanol as a function of temperature and pressure. The process is typically operated at temperatures below 32°F because the absorption capacity of the methanol increases considerably with decreasing temperatures and it also keeps the scrubbing-agent losses to a minimum. In addition, the Rectisol process is particularly suited for separate elimination of the acid

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PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY SHEET NO. _____ OF _____

SUBJECT RECTISOL ACID GAS REMOVAL SYSTEM - FIGURE 5-1

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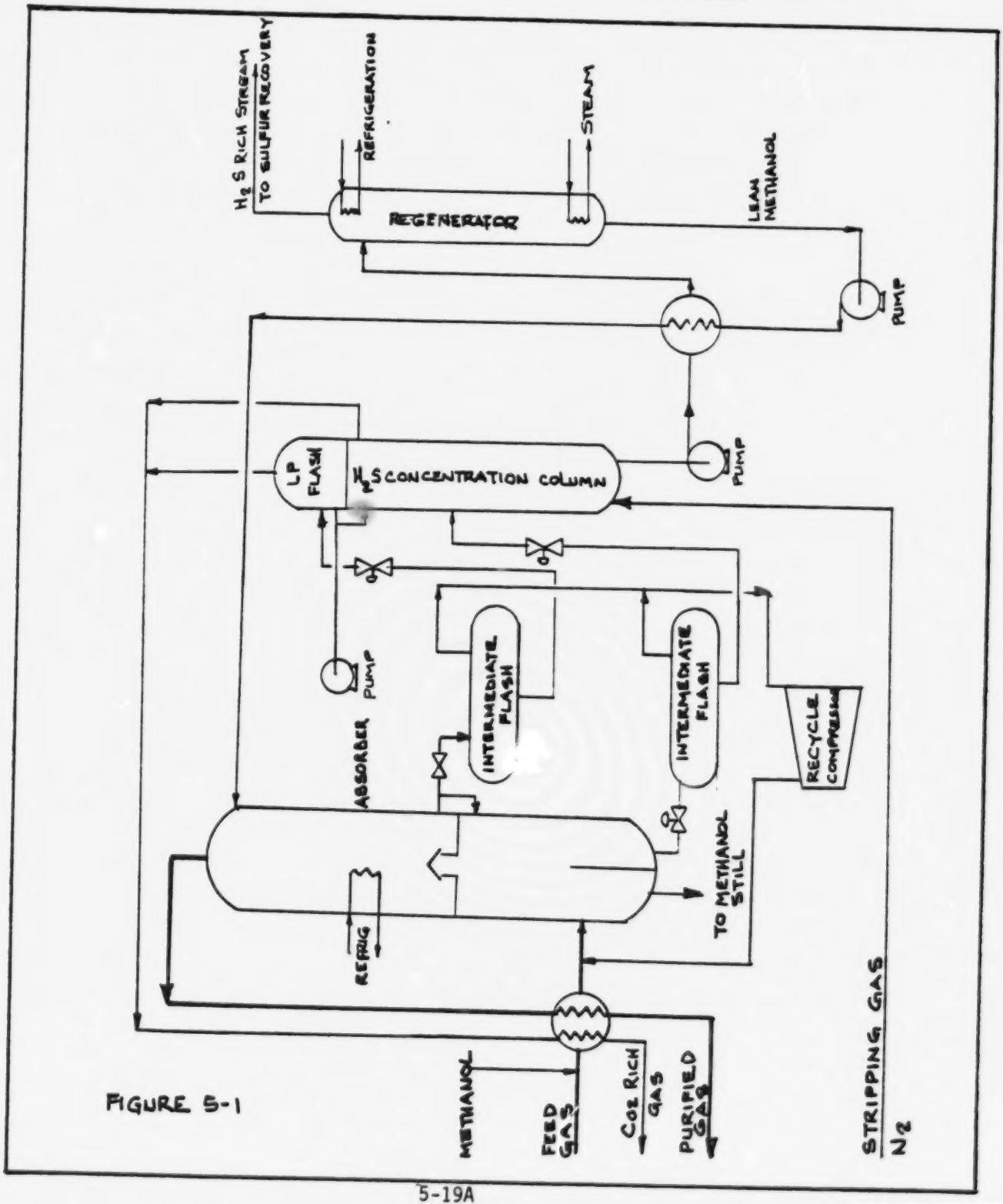

FIGURE 5-1

5.2.2.1 Rectisol Process (Cont'd)

gas fractions CO_2 and H_2S/COS due to the differing solubility of these components in methanol. This selectivity (of sulphur compounds over CO_2) enables a sulphur-free CO_2-rich gas to be obtained which may be vented directly to the atmosphere or used for CO_2 production. The Rectisol system is capable of producing a treated gas stream containing less than 0.10 ppm (Volume %) of sulphur compounds (COS + H_2S).

A typical Rectisol Acid Gas Removal System is presented on the attached schematic diagram, Figure 5-1. Entering feed gas is chilled prior to entering the Absorber by interchange with the purified product gas and the CO_2-rich tail gas. Provisions are made for methanol injection to prevent freezing water plugging in the heat exchanger. The cooled feed gas enters the bottom of the Absorber and is scrubbed countercurrently with methanol to remove the acid components. The sulphur compounds, H_2S and COS, are scrubbed in the lower section of the Absorber while bulk CO_2 removal is effected in the middle section. Final CO_2 removal occurs in the Absorber's upper section by means of lean (CO_2 and H_2S-free) methanol.

Refrigeration is used to remova the heat generated by the physical absorption process. The purified gas stream is heated by cooling the feed gas and leaves the system.

The rich methanol streams exit the Absorber and are individually flashed to an intermediate pressure level. The flashed gases (mainly CO and H_2) are recycled back to the Absorber feed gas stream via a recycle compressor. The CO_2-loaded methanol originating from the middle section of the Absorber is subsequently expanded to low pressure in the top portion of the H_2S Concentration Column thereby

5.2.2.1 Rectisol Process (Cont'd)

flashing off part of the absorbed CO_2. The bulk of the degassed methanol is returned to the middle of the Absorber. The H_2S-rich methanol originating from the bottom of the Absorber is stripped of CO_2 by nitrogen in the lower section of the Concentration Column. Washing the nitrogen and stripped CO_2 mixture with CO_2-saturated methanol in the top of the Column prevents loss of H_2S from top of the Column. Sulphur-free tail gas is withdrawn from the top of the Column and a concentrated H_2S-methanol stream is produced in the bottom. This stream is fed to the Regenerator after exchanging heat with the lean methanol from the Regenerator. The Regenerator produces a lean methanol solution which after cooling is rerouted to the top of the Absorber, and an H_2S-rich fraction from the top of the Column. This fraction is delivered to a Claus unit for the recovery of elemental sulphur.

The Rectisol process is a versatile acid gas removal system which draws its principal advantage from the selective adsorption of the sulphur bearing compounds from a feed gas stream. The system described above is one variation of the process which may be modified and adapted to fit any number of processing requirements. In order to protect the integrity of the methanol solvent pretreatment of the feed gas may be necessary to limit the introduction of condensible hydrocarbons, water vapor, ammonia, hydrogen cyanide, sulphur and nitrogen oxides, and free oxygen into the Rectisol System itself. The need for such pretreatment and its extent is, of course, also dependent upon the coal gasified and the conditions of gasification. The considerable experience achieved with the Rectisol system, including experience with those coal-derived gases, has permitted the licensors to identify and place tolerable limits on the contaminants.

5.2.2.2 Selexol Process

The Selexol Process is a proprietary acid gas removal system which was developed by the Allied Corporation in the early 1970's and is currently licensed by the Norton Co. The process is based on the physical absorption of H_2S, COS and CO_2 by the Selexol solvent (trade name for the dimethylether of polyethylene glycol) and has been demonstrated commercially for natural gas treating. The process is generally considered applicable for coal-derived gases and is capable of simultaneous dehydration to pipeline specifications. As is characteristic of all physical absorption systems, the Selexol process derives its advantages from the solubility of particular gases in the Selexol solvent as a function of temperature and pressure, and its selectivity for sulphur bearing compounds. Under suitable conditions the process is capable of producing a treated gas stream containing less than 1 ppm total sulphur.

A typical Selexol Acid Gas Removal System is illustrated on the schematic diagram Figure 5-2. The sour feed gas after cooling to approximately 40°C is fed to the bottom of the Absorber. The lower section of the Absorber removes the bulk of the H_2S and COS in the feed gas by contact with CO_2-loaded Selexol solvent from the upper section. The upper section of the Absorber effects bulk CO_2 removal concurrent with final removal of the sulphur compounds (COS + H_2S) into the ppm range by contact with lean, cool Selexol solvent. Interstage cooling of the solvent may be necessary to improve the absorption capability in the lower section of the Absorber. The purified gas exits the top of the Absorber and leaves the system.

The rich Selexol solvent exits the Absorber and is let down through a hydraulic expander which supplies part of the power requirement of the lean solution pump. It then flows to a flash drum where most of

PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY SHEET NO. _____ OF _____

SUBJECT SELEXOL ACID-GAS REMOVAL SYSTEM - FIGURE 5-2

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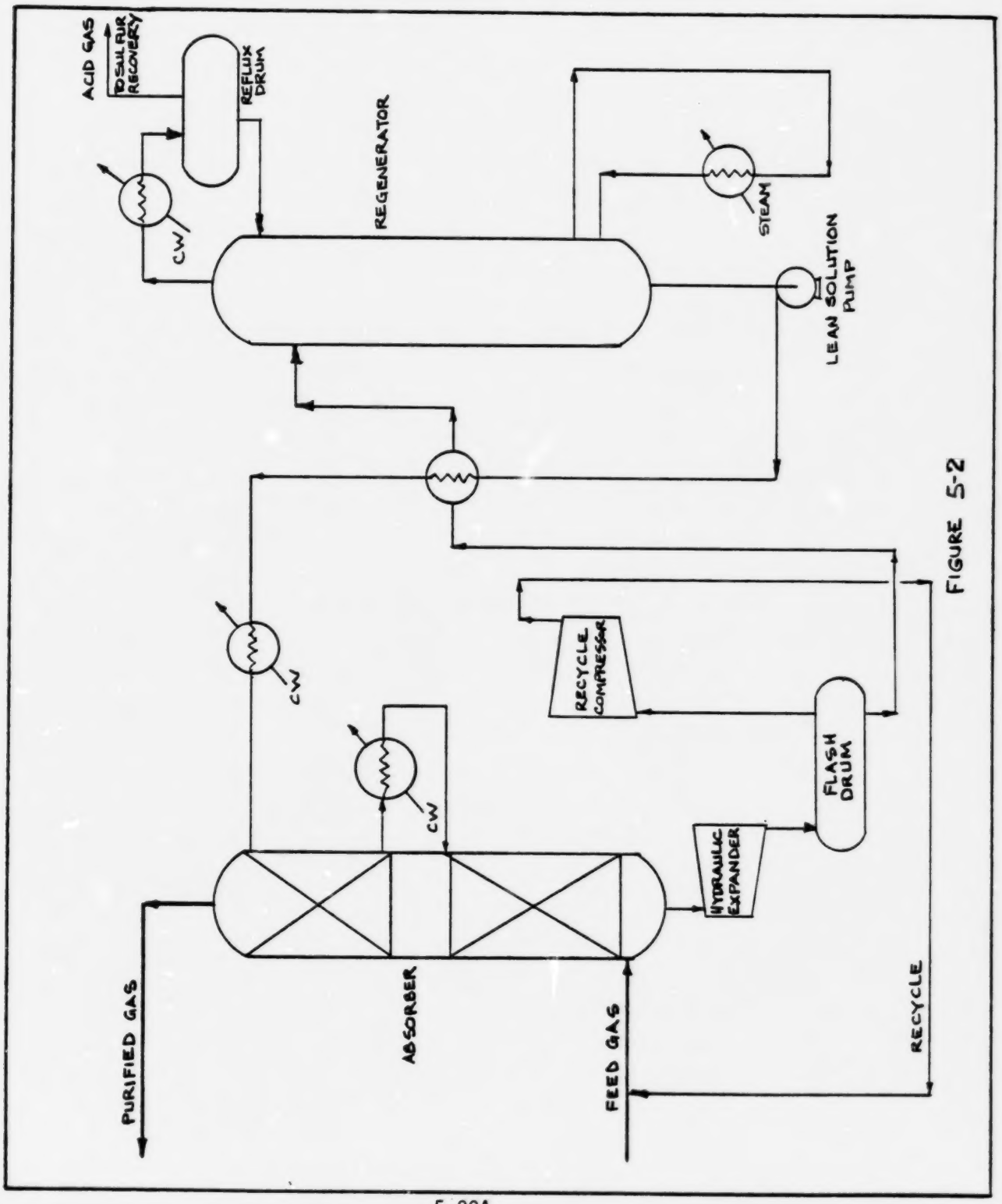

FIGURE 5-2

5.2.2.2 Selexol Process (Cont'd)

the dissolved gases in the solvent flash off. Almost all of the dissolved H_2S and most of the COS are, however, retained in the solvent because of their selective absorption in the solvent. The flashed gases are normally recycled by a compressor to the inlet of the Absorber but in some cases, they are utilized as a fuel gas. The rich solvent from the flash drum which contains the major fraction of H_2S/COS is preheated with lean Selexol solvent and fed to the Regenerator. The Regenerator strips the solvent of acid gases producing an overhead acid gas fraction for the sulphur recovery unit and a lean Selexol solvent bottoms product. The Selexol is cooled and recycled to the top of the Absorber by a lean solution pump.

The Selexol Process is a versatile one which can be modified and adapted to best suit the overall requirements for the system. While low solvent temperatures favor the solubility of the gases in the solvent, and refrigeration is therefore often used, its use is not necessary for purposes of minimizing loss of the solvent as vapor. The relatively low solubility of COS may require, if sulphur removal requirements are very stringent, the catalytic hydrolysis of the COS may require, if sulphur removal requirements are very stringent, the catalytic hydrolysis of the COS before its removal as H_2S. Both HCN and NH_3 are very soluble in the solvent and do not harm it but pretreatment to remove them is probably preferred so as to avoid problems in the downstream sulphur recovery unit.

5.2.2.3 Benfield HiPure Process

The Benfield HiPure Process is a chemisorption system for the removal of H_2S and CO_2 by a proprietary solution of hot potassium carbonate. This system, presently licensed by Union Carbide, was developed by

5.2.2.3 Benfield HiPure Process (Cont'd)

The Benfield Corporation and has over 500 units operating worldwide. Applications of the process have been used for natural gas sweetening and for purification of substitute natural gas, reformed gas and partial oxidation gases. Benfield system operation is dependent on the chemical reaction of acid gas components (H_2S, COS, CO_2, HCN, etc.) with a solution of 25 to 35% potassium carbonate with amine activators. Differences in the relative absorption rates of various components allow for the design of selective absorption (H_2S over CO_2) systems. Removal of COS and C_2S does not occur through direct absorption, but is effected through hydrolysis eventually to H_2S and CO_2 which is subsequently absorbed by the solution. The COS hydrolysis reaction rate is relatively slow making COS removal to low levels difficult, although commercial Benfield processes are reported to have achieved 75% to 99% COS conversion. Purified product gas containing approximately 10 ppm total sulphur and an H_2S rich gas suitable as a Claus Plant feed gas are achievable by staging the process.

Several variations of the Benfield process have been employed commercially, differing by the degree and selectivity of H_2S and/or CO_2 removal required. A typical non-selective Benfield HiPure process is presented in Figure 5-3. Gas feed enters the HiPure process at the bottom of the Absorber and flow countercurrently to solutions which enter at the top and at the middle of the column. The purified gas leaves the top of the Absorber and exits the system.

The contacting absorption solutions entering the Absorber differ in both temperature and composition. The solution entering the top section is cooled and is more thoroughly stripped of acid gases than

5.2.2.3 Benfield HiPure Process (Cont'd)

the lower section solution feed, and is responsible for reducing acid gases (principally H_2S) into the ppm range. The lower section effects bulk acid gas removal and hydrolysis. The two absorbing solutions operate in two separate circuits, having different temperatures, compositions and flow rates, which allow kinetic and equilibria differences between CO_2 and H_2S absorption to be used to effect the desired H_2S removal. Circulation rates and contact times are adjusted to minimize CO_2 removal if desired. The rich solutions are transported separately to the Regenerator where they are stripped of acid gases. A single concentated acid gas stream is produced in the Regenerator in the case of the non-selective design.

Ammonia, hydrogen cyanide, and sulphur dioxide are all readily absorbed by hot carbonate solutions. It is reported, however, that after absorption these compounds react further forming eventually a variety of compounds such as sulphates, thiosulphates, thiocyanates, polysulphides, and elemental sulphur which accumulate in the solution. Pretreatment of the feed gas to remove such contaminants is probably preferred to purging or blowing down of the system when handling coal-derived gases. The active components of the solvent, potassium carbonate, is very stable and not subject to degradation of loss as a vapor. The highly alkaline solution is highly corrosive and requires use of suitable inhibitors. The purified gas is essentially saturated with water vapor leaving the system. This is advantageous when the gas is used as a fuel because of its inhibiting effect on NO_x formation.

PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY SHEET NO. _____ OF _____

SUBJECT BENFIELD HI PURE PROCESS - FIGURE 5-3

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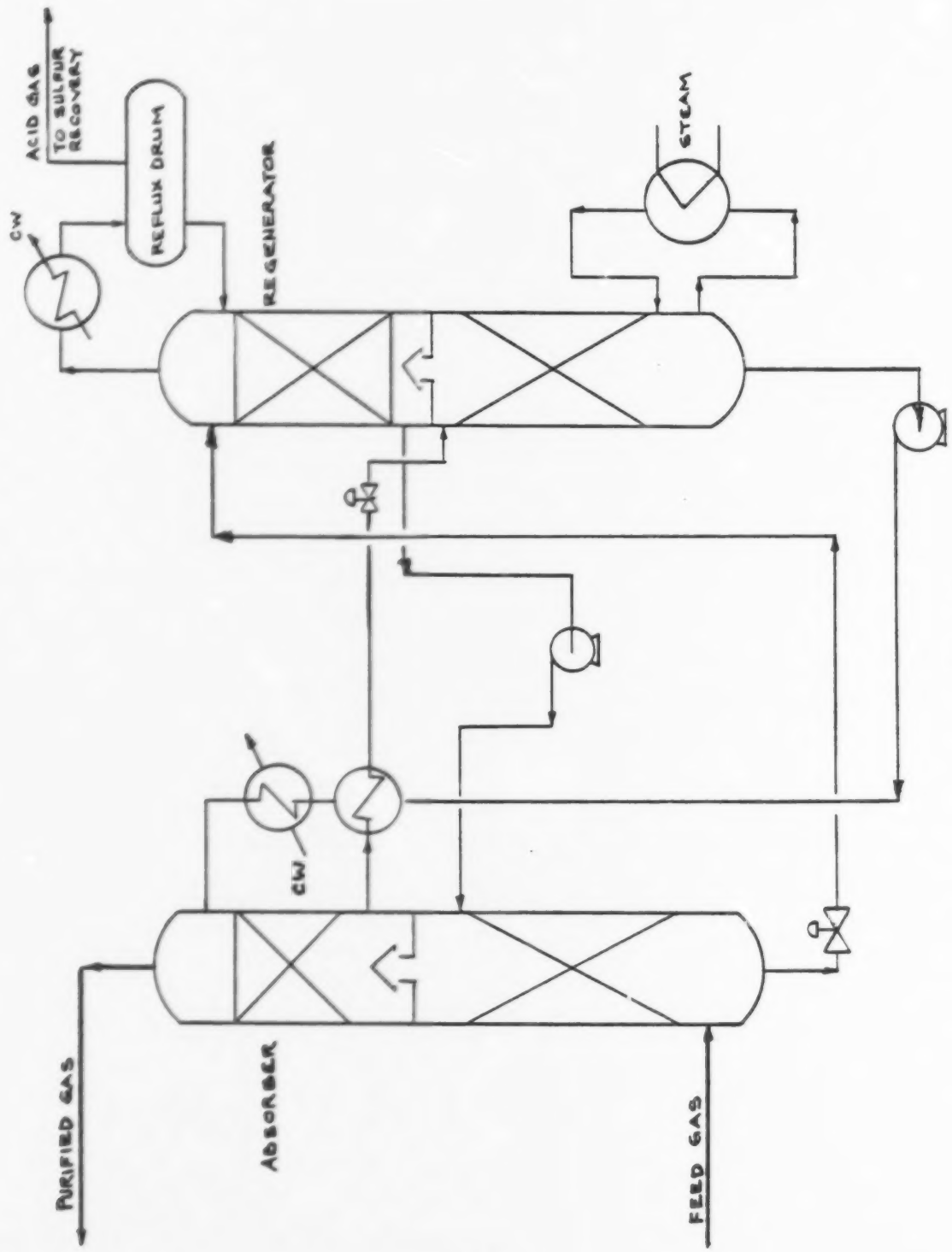

FIGURE 5-3

5.2.2.4 Fluor Econamine Process

The Fluor Econamine process is a chemical absorption system developed by Fluor Engineers and Constructors, Inc. for the removal of acid gas impurities from a feed gas stream. The absorbing medium for this process is a relatively concentrated aqueous solution of diglycolamine (DGA). Commercial applications of the Fluor process have been for H_2S and CO_2 removal from natural gas, synthesis gas and refinery gas streams. The solvent is in many respects similar to other amine solutions (MEA, DEA etc.) except that its low vapor pressure permits its use in relatively high concentrations, typically 40 to 60 percent. This results in appreciable lower circulation rates (smaller equipment sizes) and less steam consumption (operating costs) when compared to other amine solutions. Other advantages include low solvent losses (due to high vapor pressure), low absorption of heavier hydrocarbons, and suitability for processing gases containing COS and CS_2. The process's principal disadvantage is its non-selectivity for H_2S versus CO_2, but is capable of achieving total sulfur removal to 1 ppm in the product gas.

A typical Fluor Econamine process is illustrated in the attached schematic diagram, Figure 5-4. The feed gas is fed to the bottm of the Absorber where it flows countercurrently to the DGA solvent and is scrubbed of its acid gas impurities. Purified product gas exits the top of the Absorber and leaves the system. The loaded solution leaves the bottom of the Absorber, is flashed to low pressure, and preheated by interchange with hot, lean solution. The rich solution is subsequently introduced into the Regenerator for stripping of acid gases.

PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY SHEET NO. _____ OF _____

SUBJECT FLUOR ECONAMINE PROCESS - FIGURE 5-4

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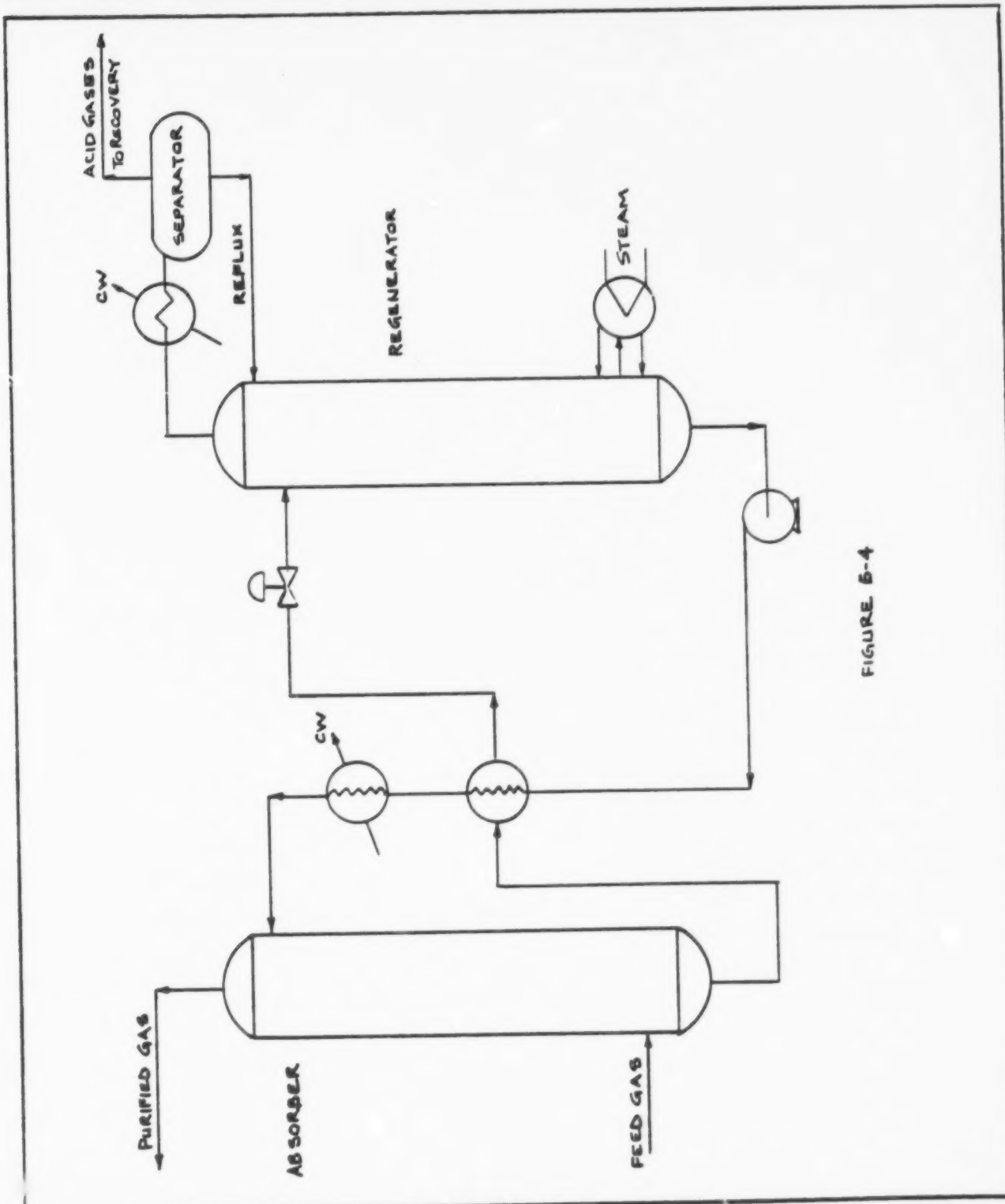


FIGURE 5-4

5.2.2.4 Fluor Econamine Process (Cont'd)

The Regenerator operates at low pressure (approximately 10 psig) and is normally steam reboiled to provide stripping. Acid gases and water vapor pass overhead to the condenser. Water is condensed and refluxed back to the tower and acid gases are directed to the plant's recovery unit. The lean solution at the Regenerator bottom is cooled and returned to the top of the Absorber.

The Fluor Econamine process has proven very effective in removing sulphur compounds to extremely low levels. Almost complete removal of COS, HCN and NH_3 can be realized. The process is reported to have higher chemical losses than a conventional MEA system. This may be due to the formation of non-regenerable products from reactions with COS, CS_2, NH_3, and HCN. Reclamation of contaminated DGA solution requires vacuum distillation because of its higher boiling point.

The major disadvantages include the relatively high cost of DGA and its corrosivity towards carbon steel. As mentioned before, the process's non-selective absorption of all acid gases may impact on selection of a suitable sulphur ecovery. A Stretford method of sulphur recovery may be necessary if the feed gas stream contains a large amount of CO_2 relative to H_2S.

5.2.3 Selection of Low Temperature Sulphur Removal System

To date the processes most proposed and used for the purification of coal-derived fuel gases have involved commercially available low temperature (below 150°C) technology. It is the intent of this study to determine the effect that use of a developing high temperature

(above 400°C) technology for sulphur removal would have on the overall thermal efficiency and the cost of combined cycle electric power production from coal-derived fuel gas.

To provide a base for comparison, a low temperature technology has to be selected which will be:

- Capable of fulfilling the requirements for the product fuel gas (Section 4.3) when used in conjunction with three different mode of coal gasification.

- Unbiased with respect to any of the coal gasification technologies used.

and would not be unnecessarily prejudicial against low temperature purification.

In Section 5.2, many available commercial processes for low temperature sulphur removal were overviewed. These processes were grouped and characterized by the type solvent used. Both the amine and hot carbonate type solvents are dependent on suitably reversible chemical reactions with acid gases for their use. The use of the organic physical solvents on the other hand, are dependent on the reversible preferential physical absorption of the acid gases. Represenatives processes using these solvent types were further reviewed so as to better examine their application. This choice of representative processes should not be constructed as indicative of the superiority of the particular system for sulphur removal. For

purposes of this study any of the processes would have been adequate and none could be expected to be optimum for use with all the gasification technologies involved.

An amine solvent system, such as the Fluor Econamine DGA Process, is not deemed a suitable choice for purposes of this study. The amine processes exhibit poor selectivity for sulphur compounds with consequently appreciable CO_2 losses from the product fuel gas. This would impose an unnecessary penalty on the combined cycle plant by increasing the net plant heat rate thereby decreasing its efficiency. The relatively high utility requirements of an amine process and the corrosivity of amines are added drawbacks. Finally, non-selective acid gas adsorption would produce an off-gas stream which might not be suitable as a feed to a cost-effective Claus sulphur recovery unit for all the cases under study.

A hot carbonate system, such as the Benfield Process, might be considered a reasonable choice for the needs of this study. The selectivity of the solvent is markedly superior to that of the amines but is not sufficiently so to permit universal application of the non-selective Hi-Pure process to all the gasification technologies of the study.

The lowest gasification temperature for the moving bed Mode of gasification produces a crude gas with a relatively high CO_2 to H_2S ratio compared to the highest temperature or entrained bed gasification mode. The flexibility of the Benfield process does not permit design of "selective" units but the added cost of such units would unduly penalize the gasification modes not requiring it. The

ability of hot carbonate to handle COS by its hydrolysis is often a desirable feature but is not necessary to achieve the degree of sulphur removal required for the study.

The qualitative review and consideration given the available processes indicated that a logical choice for the low temperature sulphur removal process of this study would be a physical solvent such as is used by the Rectisol and Selexol processes. Selexol was selected not because it can necessarily be proven the superior of the two physical solvent processes reviewed but because:

- Without refrigeration it can achieve the desired level of sulphur removal. Rectisol, while not requiring refrigeration to achieve the sulphur specification, does need it to maintain solvent loss at nominal levels.

- It is sufficiently selective to produce with a simple flow scheme, a suitable off-gas for use in a Claus sulphur removal unit in all cases. Rectisol is not quite as selective, and with gases from the low temperature gasification might require a less than simple flow scheme.

- Removal of significant quantities of COS is not necessary to produce the desired total sulphur level.

- Selexol's low vapour pressure and high flash point tend to make solvent storage and handling simpler and less hazardous in comparison to methanol.

5.0 REVIEW OF LOW TEMPERATURE CLEAN-UP TECHNOLOGY

5.3 Low Temperature 'Clean-Up' Technology for Coal Gasification

The composition of gas exiting the three types of coal gasifier covers a wide range of gaseous products and particulate, due to the gasifier characteristics. This can be attributed to the differing coal feed size, mixing behaviours, temperature, pressure and throughput that influence the design of a "clean-up" technology for the gasification process. Conversely an understanding of the gasification conditions is necessary for an effective clean-up system.

The clean-up criteria for the impurities depends largely on meeting the environmental standards and satisfying particular process requirements. Gas turbine manufacturers specifications are generally more stringent than environmental standards. Contaminate levels in the coal gas fuels must be closely maintained in order that particulates do not exceed six microns in size with the average particle size not exceeding three microns. Trace sulphur compounds in the fuel gas form sodium and potassium sulphates in the gas turbine and cause corrosion-erosion and deposition, and affect blade life over 649°C (1200°F). Tar compounds also have to be removed from the gas prior to use in a combustion system to avoid condensation problems in the burners and distribution mains.

Westinghouse Electric Corporation in the EPRI funded "High-Reliability Gas Turbine Combined-Cycle Development" study have stated the following limits on coal fuel gas contaminants (Ref. 1).

5.0 REVIEW OF LOW TEMPERATURE CLEAN-UP TECHNOLOGY

5.3 Low Temperature 'Clean-up' Technology for Coal Gasification

Fuel Gas Contaminants to the Combustor (PPMV)							Fuel Gas Particulate
	H_2S	HCl	KCl	NaCl	NH_3	COS	(Grains/Scf)
Cold-Gas Clean-up	60	.003	.01	.004	6	144	.06

SECTION 5 - CONCLUSIONS AND RECOMMENDATIONS

Selexol was selected as the most suitable low temperature sulphur removal system because of its advantages in storage and handling and its effectiveness in sulphur removal.

6. ASSESSMENT OF HIGH TEMPERATURE GAS PURIFICATION

6.1 High Temperature Particulate Removal Technology

6.1.1 Factors Considered in Technology Evaluation

In the fuel gas leaving the coal gasifier, fractions of various constituents of the coal such as unburned carbon and ash can be entrained and carried over to the sulphur removal and combustion turbine units. Dependent upon the specific gasifier and operating conditions the particulate loadings may be sufficiently high to require gas clean-up prior to sulphur removal and/or combustion turbine equipment to achieve acceptable life and performance. Alkali metals in different physical forms (vapour, solid) may also be present at levels that require reduction prior to gas turbine combustion.

The plugging of the high temperature sulphur removal vessels with particulate material entrained from the gasifier is a potential problem. Particulate removal facilities installed between the gasifier and the sulphur absorption vessels must protect against the plugging of the sorbent bed in operation when the loading of particulate in the fuel gas is normal and in upset periods when the fuel may carry abnormally high loadings of dust.

Downstream of the sulphur absorbent bed it may be necessary to have some form of polishing filter to protect the gas turbines. The principal concern is the potential presence of entrained alkali metals in the fuel gas which can cause corrosion and erosion of turbine blading.

The high temperature and pressure particulate removal system to be considered applicable must satisfy to some extent the following criteria.

- Capable of reducing the particulate matter that exists from the three gasification systems which are consistent with the product gas specifications established in Section 4.3. This includes both the mass load removal and particle size requirements.

- Capable of reducing the particle loadings that exit from the three gasification processes to avoid plugging of the sulphur absorption vessels.

- Operate efficiently and effectively at a temperature and pressure which is consistent with the sulphur removal and combustion turbine system.

- The process should be practical and simple, and be in both the filtration and cleaning cycles.

- The process should not inflict high pressure drops in the filtration cycle.

- The process should not require exotic materials of construction for operation in a high temperature and pressure environment.

While the capital and operating costs of the particulate removal system will effect the production cost of power, they should not be considered a prime factor in the evaluation of the alternate particulate technologies. It is of greater importance to select the system based on its ability to fulfill the overall process specifications.

6.1.2 Review of Technology

6.1.2.1 Cyclones

Cyclone dust collectors rarely meet air pollution control code requirements used alone, and therefore should be evaluated as a precleaner, followed by other dust collectors. Using a cyclone precleaner will not necessarily reduce the size of the secondary dust collector, but can often both simplify the design and increase the reliability of the secondary collector and accessory equipment. In conventional applications, cyclones have been evaluated with the following secondary collectors to meet emission control regulations:

- Electrostatic Precipitator
- Wet Scrubber
- Baghouse
- Cyclones in Series

Generally, cyclones have been limited to the collection of particulate matter 10 microns in diameter and larger. The overall and grade efficiency curves for cyclones are primarily affected by the density of particles, gas density, temperature of the gas stream, gas inlet velocity and cyclone diameter.

Recently, Exxon Research and Engineering Company, has evaluated the performance of cyclone separators operating at high temperatures (700-900°C) and pressure 1000 KPa. The tests have been conducted on flue gas from a pressurized fluidized bed combustor. The flue gas exiting the combustor passes through three cyclone separators. The solids collected in the first cyclone are returned to the combustor with the collected solids from the second and third cyclones removed through lockhoppers.

The results show that the particulate captured by the second cyclone had a mass median particle size of 15-20 microns with an overall collection efficiency of about 95 percent. Particulate captured by the third cyclone had a mass median size of 3 to 5 microns and an overall collection efficiency of about 90 percent. Particulate concentration in the flue gas leaving the third cyclone was generally 0.04 to 0.7 g/Sm^3 with a mass median size of particulate ranging from 0.7 to 3 microns. These results are summarized in Table 6.1-1.

To enhance the collection efficiencies of conventional cyclones, recent research efforts have focused on acoustic augumentation. The acoustic agglomerator uses the oscillating flow of an acoustic field to cause particle collision and agglomeration. The resulting large particles are then more readily separable by conventional methods. Acoustic agglomeration (AA) could be applied effectively for the removal of airborne dust particles when used as a preconditioner in series with conventional cyclones. The dust laden gas

would first enter the AA chamber. As it passes through the chamber, the gas is subjected to the action of sound waves radiated by a sound generator, which can be a mechanical vibrating piston, siren or other device. In the AA chamber, smaller particles (diameter <10 micron) agglomerate into larger ones which could be effectively removed by conventional cyclones.

TABLE 6.1-1

PARTICULATE CONCENTRATION AND SIZE

Cyclone	Particle Concentration (g/Sm^3) Passing	Mass Median Size (Microns) Captured	Mass Median Size (Microns) Passing	Cyclone Efficiency
First	10	-	20-25	-
Middle	0.4-0.6	20-25	3-5	95
Final	0.04-0.07	3-5	0.7-3	90

6.1.2.2 Electrostatic Precipitators

Since they were introduced to the utility industry in the 1920's, electrostatic precipitators (ESP) have become the most widely used means of controlling fly ash emissions from coal fired boilers. In practice two general types of ESP's have been used commercially, namely, hot and cold side systems, which typically operate at inlet temperatures of 400°C (752°F) and 150°C (302°F) respectively. To date ESP's have operated at approximately atmospheric pressure.

There are three fundamental steps in the ESP process:

- Charging the particles suspended in the gas stream
- Collecting the charged particles
- Removal of the collected particles into an external receptacle (hopper).

Charging is accomplished by applying a high dc voltage to an electrode system. The pressure of a grounded electrode near the charged electrode gives rise to

the formation of a corona and a unidirectional field. Particles travelling between the two electrodes acquire charge from the corona discharge and are driven toward the collecting electrode by the electric field. All commercial precipitators use negative polarity. Particles remain on the collecting electrode, or plate, until they are dislodged by a mechanical blow, rapping, and fall by gravity into an external vessel usually some form of hopper.

In controlling fly ash emissions from coal fired boilers by the use of precipitators it is known that the electrical resistance of the particulate matter, resistivity, is an important factor in their design and operation. Resistivity is a function of not only the chemical composition of the ash but also the flue gas temperature, the amount of unburned carbon and the concentration of certain flue gas constituents such as sulphur dioxide and water vapour.

It has been demonstrated that trace quantities of sulphur dioxide in the gas play a critical role in the fixation of moisture on the dust surface, providing increased surface conductivity. Usually in order for resistivity to be below the critical value for a normal air heater outlet temperature (approx. 150°C) there must be at least 9 to 12 ppm of SO_2 in the flue gas which corresponds to approximately 1.5 to 2 percent sulphur.

This effect is modified by the effect of sodium in ash when firing western lignite and western sub-bituminous coals. These fuels are low in sulphur and therefore would be expected to have high resistivity at 150°C. However, some contain up to 10 percent Na_2O in the ash which in the combustion process is largely volatilized. Under there conditions, Na_2O would be expected to react with available SO_3 to form Na_2SO_4 so that, upon reduction of temperature, condensation and absorption are believed to occur, resulting in a surface enrichment of Na_2SO_4 on the alkaline (high CaO + MgO) ash. Na_2SO_4 is a fairly low resistivity material at 150°C so that the overall resistivity of the ash is low. Ash resistivity can be correlated with its constituents by the ratio:

$$\frac{CaO + MgO}{Na_2O + SO_3}$$

This resistivity increases with increase of the ratio.

CaO exhibits high resistivity in the 150°C range, even at 20 percent moisture in the gas phase, hence in the case of high sodium fuels, a surface conductivity mechanism is apparently provided by the Na_2SO_4 even though the ash is highly alkaline and would ordinarily be high in resistivity.

The dust resistivity is a measure of how the dust will collect. For particles having a mass resistivity of less than 10^6 ohm-cm, the particles are relatively conductive and therefore tend to re-entrain on reaching the receiving electrodes where they lose their charge and are repelled back into the gas stream.

At the other end of the scale with particles having a resistivity greater than 10^{12} ohm-cm the precipitation process is adversely affected due to the layer of dust on the receiving electrodes, which can act in one of two ways. The condition of reverse or back ionisation can exist in which discharge takes place from the receiving electrode giving abnormally low electrode voltages and high currents, with the reverse flow of corona reducing the precipitation efficiency. The other effect is that the dust layer acts as a resistance in series with the precipitator and the electrode voltages appears at a normal level but the current is abnormally low. The net effect in both cases is to give a serious reduction in the effective migration velocity, and hence lowers the efficiency of the precipitation plant.

A typical curve for ash resistivity versus temperature can be generated with high and low precipitation resistivity limits (range of precipitator capability). It is apparent from the ash resistivity curve that the resistivity decreases toward the low limit of precipitator capability at the hot precipitator temperature 400°C (752°F). Higher temperatures than this may decrease the ash resistivity below the low limit of precipitator capability, and therefore, it must be assessed for each type of coal.

High resistivity affects ESP operation especially if the layer of ash is allowed to build up to the collection of plates. Because of the resistance, a

voltage drop develops between the outer layer of the ash and the grounded metal collecting plate. If high enough this voltage drop makes removal of the collected dust very difficult because of the resultant strong electrostatic forces which hold the dust to the plate. Heavy rapping (with potential re-entrainment losses) must be employed to dislodge the collected fly ash.

Low resistivity of the particulate matter can affect the collection efficiency of an ESP. The resistivity of the particulate matter, among other factors, is generally decreased by increases in carbon content. For example, it has been experimentally demonstrated that ESP's are capable of removing 95-97 percent mass load removal up to 825°C, while carbon containing particles remain entrained.

ESP's have a non-linear fractional efficiency, that is, the efficiency changes with the size of the particle to be collected. Generally, there is a noticeable drop in efficiency for particles below 5 microns.

Research - Cottrell has recently reported on the feasibility of ESP's at temperatures and pressures varying from ambient conditions to approximately 1100°C and 3550 KPa, respectively. It has been demonstrated in bench scale testing that a wire-pipe electrode system can produce a stable corona discharge at all temperatures evaluated subject to appropriate choices of electrode dimension, polarity and pressure. In addition, it was revealed that as temperature and pressure are increased together precipitation is possible at significantly higher voltage than at normal conditions. This is an important factor when assessing the viability of ESP for high temperature particulate removal applications because the rate of particle collection is roughly proportional to the square of the electric field strength within the precipitator.

Therefore, the net effect of higher operating voltages is either a proportionally larger increase in particle collection efficiency or a corresponding decrease in the required precipitator size for a specific efficiency.

6.1.2.3 Ceramic Fiber Filters

Westinghouse has been investigating woven and felted ceramic fiber filters for hot gas particulate removal of flue gases from pressurized fluidized bed combustion power generation. At Westinghouse's Waltz Mill Hot Gas Clean-up Tests Facility a woven ceramic filter supplied by the 3M Company and Buell Envirotech and a fibrous (felted) filter supplied by Acruex/Aerotherm were tested.

The woven ceramic unit was sized and designed for operation at filter face velocity (volumetric gas flow per unit time per cross-sectional area of the filter) conditions typical of current low temperature bag designs, 0.6 to 1.2 m/min, and cleaned on line by employing a high pressure, short duration back pressure pulse.

The felted bag unit was sized and designed for filter face velocities much higher than conventional filter bag operation of 3.05 to 6.1 m/min. At these filter face velocities the individual bags for cleaning were taken off line (the forward flow is stopped during the cleaning cycle).

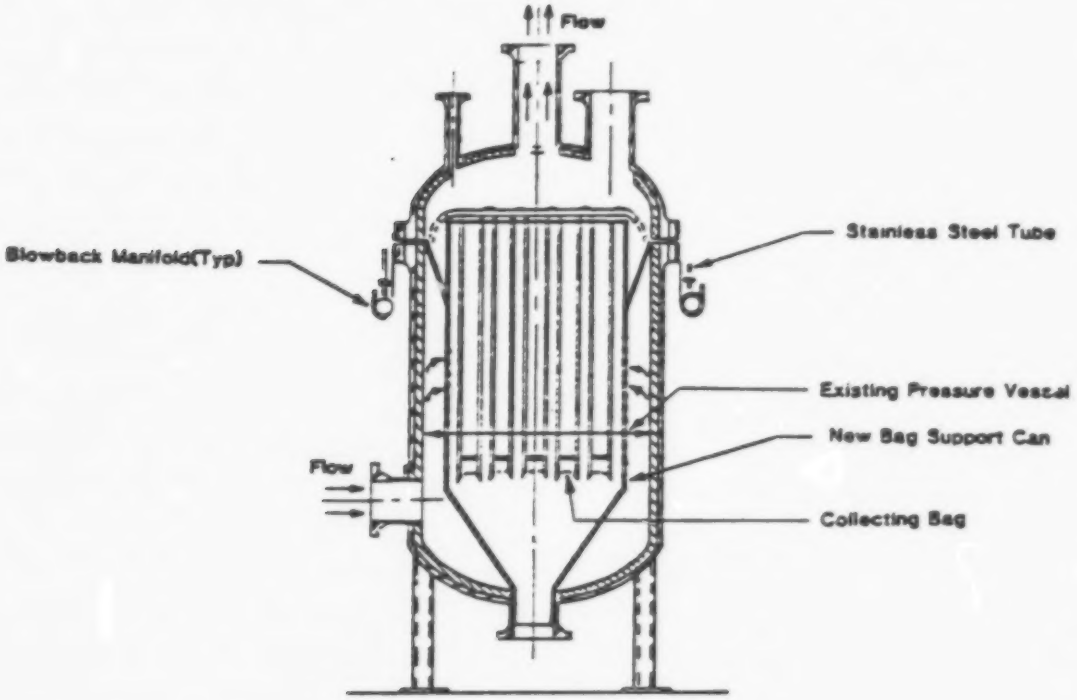
The tests were conducted using ash obtained from an Exxon pressurized fluidized bed miniplant and on a composite material consisting of 25 percent limestone plus 75 percent ash. The mean mass diameter for the composite material was approximately 15 microns, with no more than 10 percent of the dust having a size range of 2 microns or less.

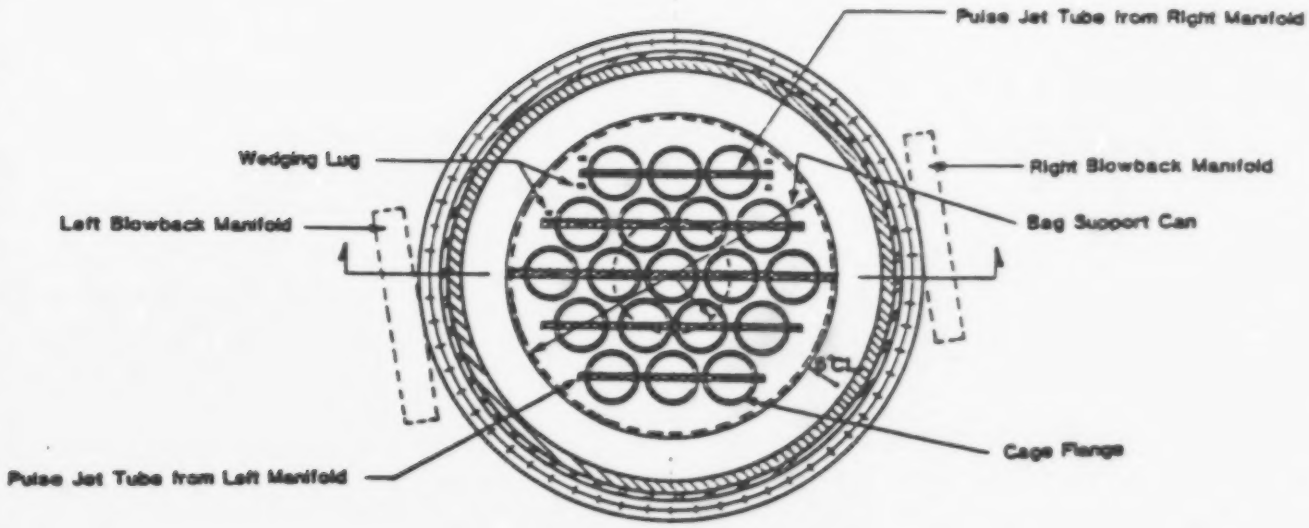
The general arrangement schematic of the woven ceramic filter unit tested is illustrated in Figure 6.1-1. The unit consists of a 19 bag array suspended in a pressure vessel. The dust laden gas is introduced through the wall of the pressure vessel, through the perforated wall of the bag canister, up the inside of the bag and perforated support cage and out the top of the pressure vessel. Each bag is cleaned by a pulse jet nozzle located at the top of the bag.

PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY SHEET NO _____ OF _____

SUBJECT WOVEN CERAMIC FILTER UNIT

FIGURE 6.1-1





WOVEN CERAMIC FILTER UNIT

6-8-A

Table 6.1-2 summarizes the overall physical and chemical characteristics of the ceramic fiber material used.

The initial testing of the filter unit correspond to a period of "bag conditioning" where the relatively large spaces between the fiber weave partially filled with dust particles to establish an identifiable filter cake.

At a pressure of 1138 KPa and temperature of 427°C, tests were conducted at two mass flow rates with filter face velocities of 0.85 m/min and 1.71 m/min. During steady-state operation (after bag conditioning), at the lower face velocity tests, and relatively high inlet dust loadings, the overall collection efficiency was 99.9 percent or greater with outlet loadings less than 2 ppm (0.0012 gr/scf). At higher face velocities of approximately 1.71 m/min the overall collection efficiency was 99.6 percent. The results showed that at the low filter face velocity and once the bag was conditioned, high efficiencies were achieved for particle sizes less than 10 microns. At the higher filter face velocity, 1.71 m/min, the grade efficiency curve showed slightly lower performance.

In the first high temperature (815°C and 1138 KPa) test conducted at a 1.71 m/min filter face velocity, the measured overall collection efficiency was 99.9 percent which is comparable to the results in the lower temperature tests. The results from the second high temperature test were divided into three segments. In the first segment, passage conditions were stable and the overall collection efficiency was approximately 99.2 percent. Further tests revealed lower collection efficiencies and it was concluded and confirmed that a failure in a bag had occurred.

TABLE 6.1-2

PHYSICAL/CHEMICAL CHARACTERISTICS OF THE CERAMIC FIBERS USED IN MANUFACTURING THE FELTED BAGS

• Composition	95% Al_2O_3 5% SiO_2
• Trace Name (Supplied by ICI)	Saffil Alumina
• Service Temperature	3000°F
• Fiber Density	3.4 g/cm^3
• Fiber Diameter	3 µm nominal
• Fiber Length	1 - 2 inch

The tests showed that the filter bags can be cleaned by jet pulse with pressure drops ranging from 0.75 KPa to 2.0 KPa. The effectiveness of the bag cleaning was found to be dependent on the pressure ratio between the bag operating conditions and the pulse jet pressure

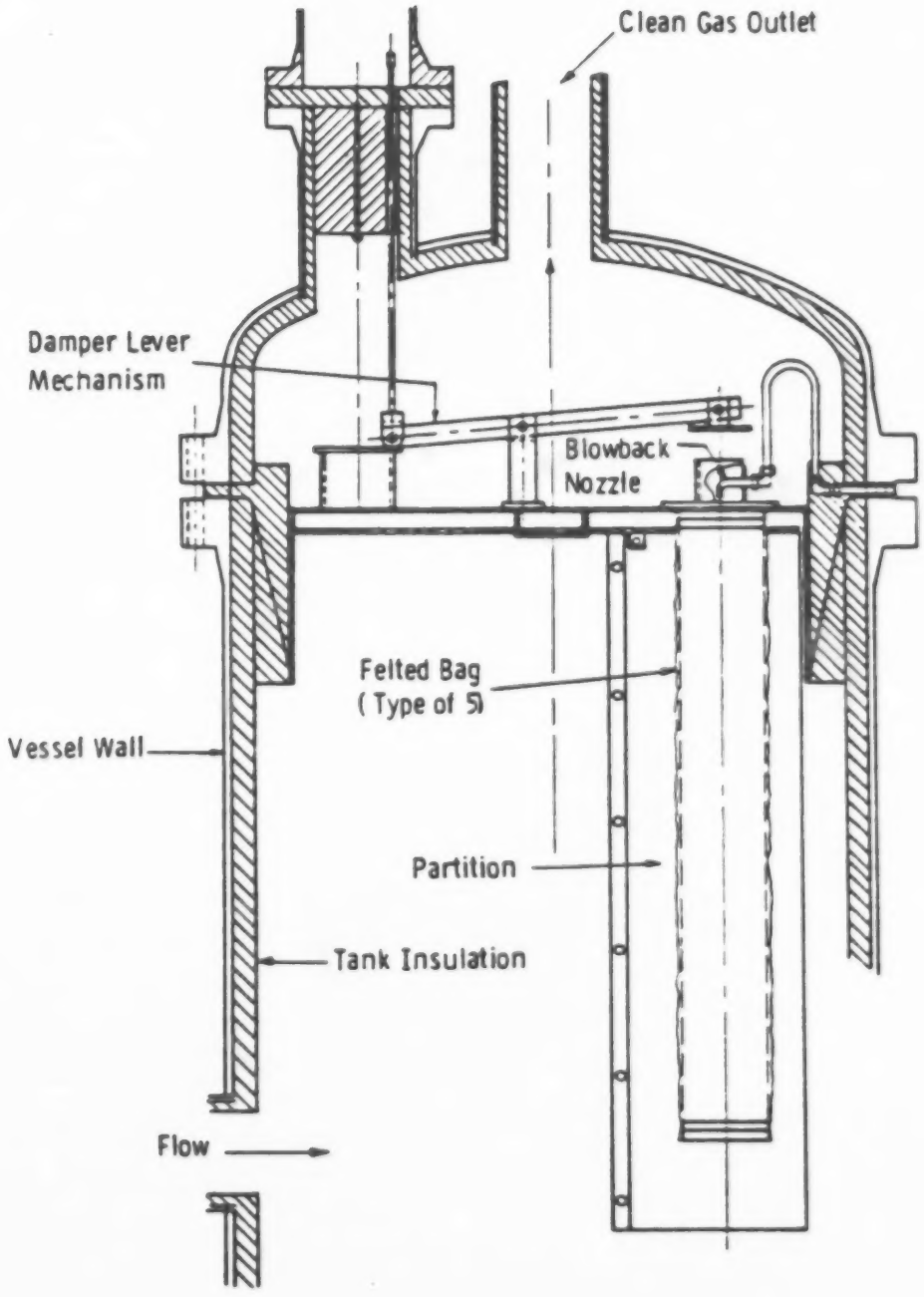
$$\frac{P_{\text{blow back}}}{P_{\text{test}}}$$

as opposed to the pressure difference, $P_{\text{blow back}} - P_{\text{test}}$.

Effective bag cleaning was achieved when the pulse jet pressure ratio was maintained at about three or above.

The general arrangement drawing of the felted bag unit is illustrated in Figure 6.1-2. The unit consisted of a 5 bag array suspended in the pressure vessel. The dust laden gas flow is introduced through the wall of the pressure vessel at the bottom nozzle and flows up around the five felted bags, through the ceramic felt and metal support cage and then up through the centre of the bags to the clean-side. A blowback nozzle is positioned at the top of the bag (clean gas discharge) and centred on the vertical axis of the bag and support cage.

FIGURE 6.1-2



FELTED CERAMIC FILTER UNIT

Table 6.1-3 summarizes the physical/chemical characteristics of the ceramic fibers used in manufacturing the felted bags.

The hot gas cleanup tests on the felted ceramic bag filter were conducted at two temperature conditions, 427°C and 815°C at a pressure of 1138 KPa. The overall collection efficiency for these tests ranged from 99.5 to 99.8 percent, with inlet dust loadings from 421 ppm (0.024 gr/scf) to 7388 ppm (4.42 gr/scf) and outlet dust loadings ranging from less than 1 to 34 ppm.

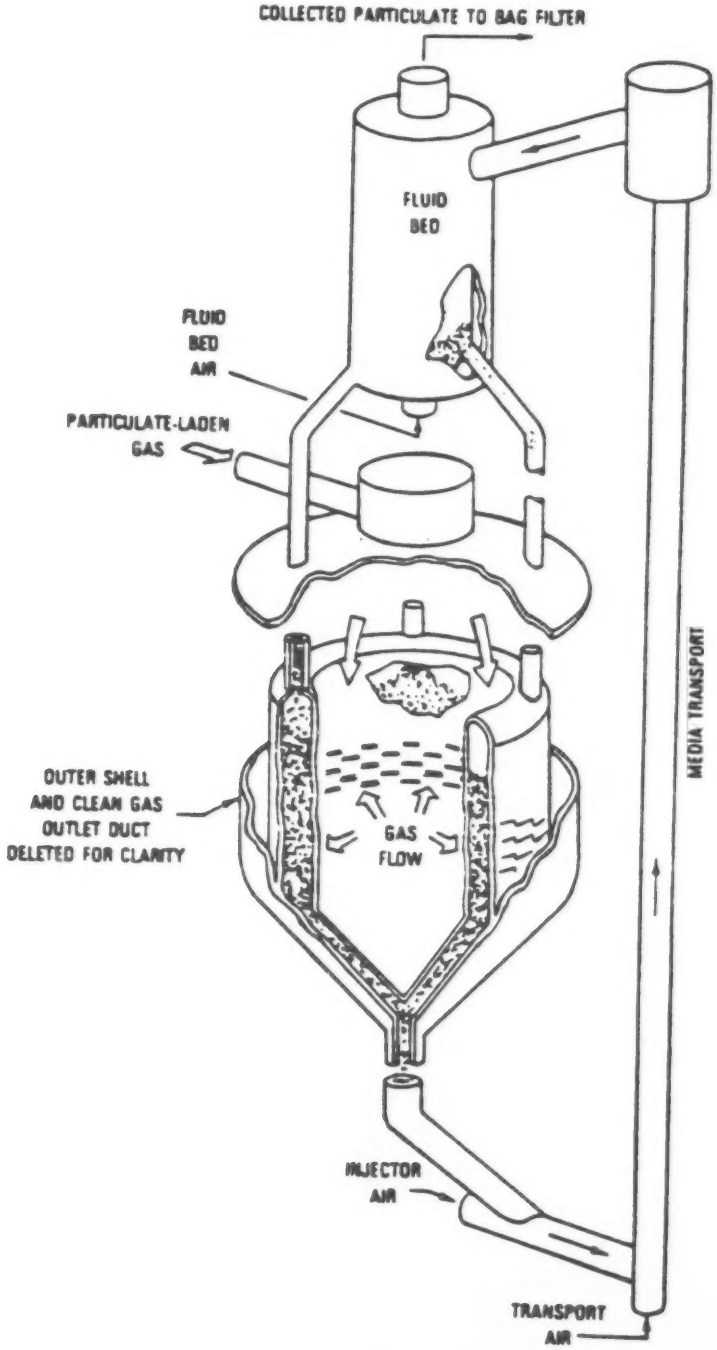
TABLE 6.1-3

PHYSICAL-CHEMICAL CHARACTERISTICS OF THE CERAMIC FIBER IN THE WOVEN (BUELL) TEST UNIT

Composite	62% alumina 14% boria 24% silica
Base Color	White
Denier* (390 filament yarn)	900
Tensile Strength	250 x 10^3 psi
Tensile Modulus of Elasticity	22 x 10^6 psi
Strand Strength (390 filament yarn)	5.0 lb
Filament Diameter	0.44 mil
Density	2.5 g/cc
Extended Use Temperature	2550°F
Short-Term Use Temperature	2900°F

The filter unit was operated over a relatively high pressure drop range, 7 to 13 KPa, compared to conventional low pressure, low temperature filters. At relatively low blowback reservoir pressures, 1379 to 1724 KPa, an increase in the base line pressure drop was observed, indicating ineffective cleaning. At higher pulse jet pressure blowback conditions, 2413 to 2758 KPa, the base line

FIGURE 6.1-3



CROSS-FLOW MOVING-BED GRANULAR FILTER
(COMBUSTION POWER COMPANY, INC.)

operating pressure drop was reduced significantly after pulsing, indicating a more effective removal of dust during blowback. This corresponds to a critical pressure ratio of approximately three for effective bag cleaning which is consistent with the results obtained with the ceramic woven bag tests.

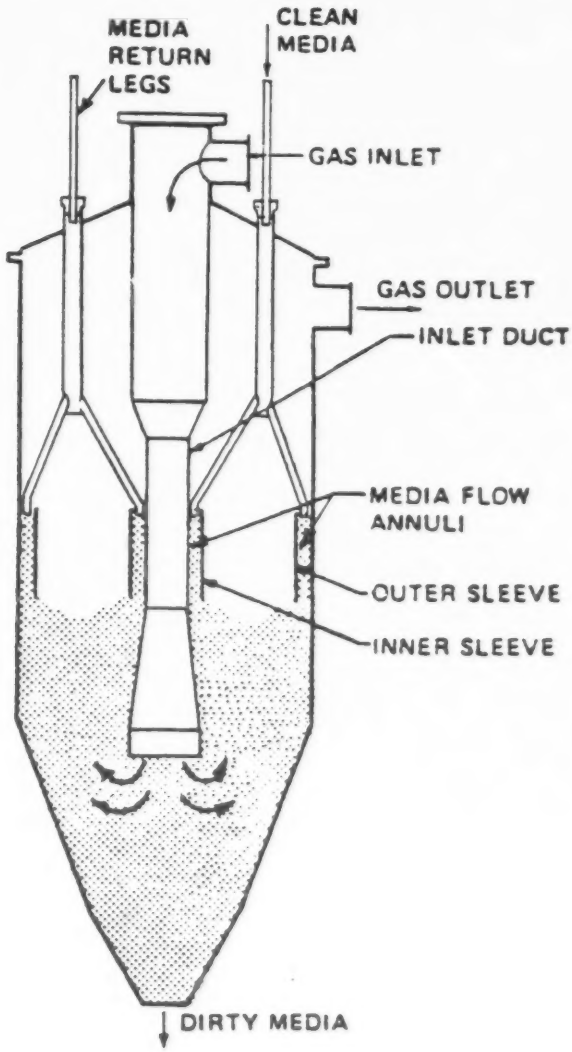
6.1.2.4 Granular Bed Filters

Granular bed filters in various modes of operation have been used for particulate removal from gas streams for many years. Currently, their exploitation in coal conversion processes for the removal of fine dust and trace metals have been pursued with varying degrees of success.

The Combustion Power Company has developed cross flow screened granular bed filters for high temperature filtration of stack gas particulate and fugitive dust cleanup. In this device, (Figure 6.1-3) the dust laden gas is allowed to flow radially outward in cross flow through an annulus of granular collecting media. The dirty media is pneumatically transported to a fluid bed located just above the filter housing and after being cleaned is cycled back into the filter. The bed filter material is of a special alumina for high temperature applications.

In initial testings, plugging problems were encountered in both high and low temperature operation. The screens of the moving granular bed filter were blinded when the flue gas was filtered for particulate matter from an atmospheric fluidized bed combustor burning Illinois Number 6 coal in dolomite. The screen blinding was attributed to the adhesive/cohesive nature of the coal ash and sulphided sorbent composite. This led to the development of a screenless countercurrent flow design for the filter as illustrated in Figure 6.1-4. In this design, no plugging problems were encountered in four test segments with continuous operation of the filter for 100, 200, 300 and 400 hour periods.

FIGURE 6.1-4



HIGH TEMPERATURE SCREENLESS COUNTER-CURRENT MOVING-BED GRANULAR FILTER (COMBUSTION POWER COMPANY, INC.)

In these tests, particulate laden flue gas exiting a primary cyclone of an atmospheric fluidized bed combustor was allowed to flow through the screenless granular bed filter at 870°C. The particulate capture efficiency remained approximately 99.0 percent for filter pressure drop values ranging between 6.0 and 7.0 KPa and an inlet gas flow velocity and particulate loading of 0.18 m^3/s and 6.0 g/m^3, respectively. Although, further testing revealed that the particulate collection efficiency decreased from 99.0 percent for particles having an aerodynamic particle diameter greater than or equal to 4 microns, to about 93.2 ± 1.0 percent for particles of about 0.3 microns.

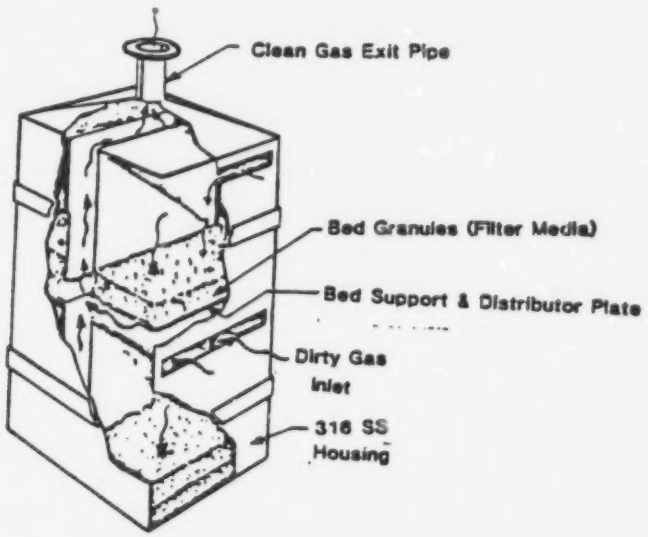
Westinghouse and Exxon have conducted tests on a fixed bed granular filter supplied by Ducon, Inc. A single element Ducon filter is illustrated in Figure 6.1-5. The unit consists of four compartments which operate in parallel both in filtration and cleaning (blowback) cycles. Each compartment contains a shallow bed of granular particles supported on an appropriate distributor plate which serves as the filter support media. In the filtration cycle, the particle laden gas enters through a slot at the top of each compartment and flows down through the filter bed. The cleaned gas from all compartments combines and exits out of the element. In the cleaning cycle, the beds are cleaned by fluidizing them with high pressure air flowing in reverse direction for a short duration. Sufficient freeboard height is made available in each bed compartment so that the media granules are not lost in the cleaning cycle while the dust particles are ejected from the bed.

Westinghouse conducted their initial tests in a bench scale low temperature and pressure system. Westinghouse obtained collection efficiencies ranging between 99.7 and 99.8 pecent, for ground limestone dust particles (100 percent < 10 microns) with a maximum pressure drop ranging between 5 and 120 KPa during a test period of 12 hours. In their high temperature (600°C), low pressure runs, the collection efficiency ranged between 96.0 to 99.9 percent.

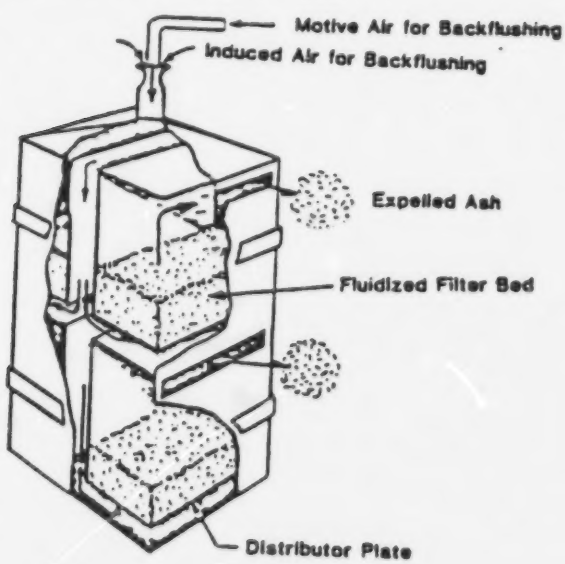
Exxon has tested a modified Ducon granular bed filter in conjunction with their mini-plant pressurized fluidized bed combustion system operating at

FIGURE 6.1-5

GBF ELEMENT - FILTRATION MODE



GBF ELEMENT - CLEANING MODE



temperatures up to 980°C and pressures up to 1000 KPa. In their tests the filter media was 300 to 600 micron quartz particles and 840 to 1400 micron alumina particles. The particulate collection efficiencies were 90 to 95 percent for outlet particulate concentrations of about 0.1 g/m^3. The particulates in the filtered gas had a mass median size of about 3 microns, with about 10 percent larger than 10 microns. Filtration velocities ranged between 20 to 24 m/s.

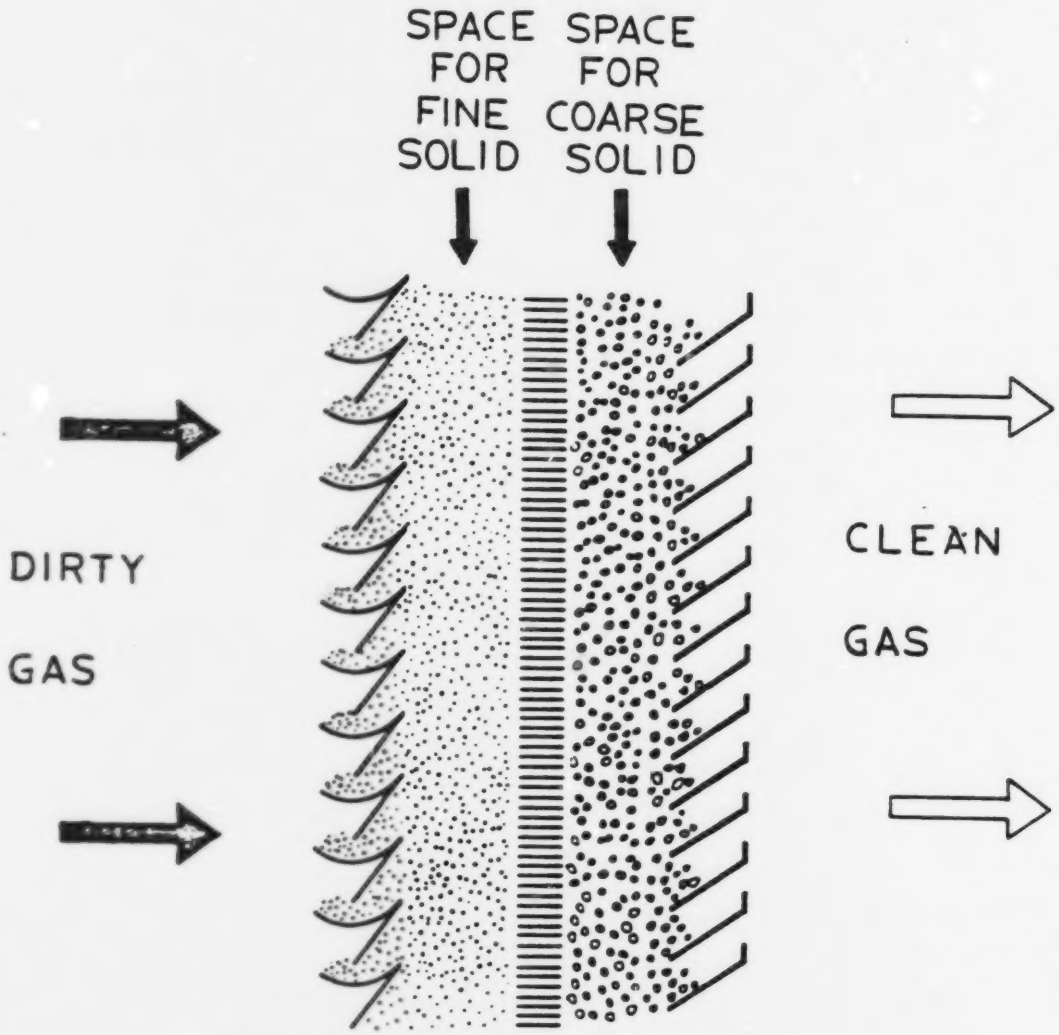
The cleaning cycle usually required blowback every 10 to 20 minutes during which time the filter pressure drop increased by about 14 KPa above its base line value. The blowback period ranged between 2 and 30 seconds, with a superficial gas velocity between 0.15 to 0.75 m/s. Dust particles were found to be returned in the filter bed (10 to 30 percent) and were uniformly distributed throughout the filter bed.

Under the direction of Professor A.M. Squires at the City College of New York (CCNY) the concept of a panel bed filter was investigated. In this device, the gas to be treated flows in a horizontal direction across a bed of granular solids supported in a tall, thin panel. The bed is supported in place by louvred walls resembling Venetian blinds rigidly fixed in an open position was illustrated in Figure 6.1-6. Dust removed from the treated gas accumulates on the granular entry surfaces of the panel betwen the louvres which comprise the walls.

For removing solids from the gas entry surface a technique termed "puff-back" is employed. It is accomplished by applying a surge of backflowing gas from the clean side of the panel. The sharp surge of gas produces a substantially uniform spill of solid from each gas entry surface when applied in reverse direction across the panel, provided that the gas surge has the proper intensity.

At CCNY, efficiencies of 99.9 percent have been achieved for removal of redispersed power station fly ash at normal stack gas loadings in atmospheric air

FIGURE 6.1-6



Cross-Section Through a Panel Bed Filter

at a temperature of approximately 150°C. Tests of a laboratory panel bed at about 538°C at the Morgantown Energy Technology Centre have shown that finer sands are required to achieve a given filtration efficiency as the temperature is increased.

For commercial application, Dr. K.C. Lee of CCNY has outlined an idea of how the panel bed may be incorporated into a unit for the filtration of the fuel gas leaving a gasifier. In the Lee design, four square panel bed elements are enclosed in a single pressure vessels as shown in Figure 6.1-7. Over the top of each gas exit nozzle from the centre gas duct of each square panel element there would be a cup-sleeve. The cup-sleeve would be moved up during filtration by a specific panel bed element and would be moved down to cover the clean gas exit nozzle when applying puff-back through the cup-sleeve.

Recent development efforts have also focused on electrically augmented granular bed filters. The Combustion Power Company has developed an electroscrubber granular filter whose schematic is illustrated in Figure 6.1-8. It consists of a cylindrical vessel containing two concentric cylindrical louvred screens. The annulus between the screens is filled with the gravel media. An electrostatic grid in the form of a cage is located in this media. A high voltage is applied to this grid and the electric field generated between this conductor and the inner and outer screens enhances the particulate collection efficiency as the dust laden gas passes through the media. It is claimed that the enhancement is due to the frictional charging of the particles as the dust laden gas flows through pipes and cyclones. The cleaned gas exits through the outer screen. The filter media moves continuously downward and is recycled into the filter after cleaning.

It is claimed that the electrostatic device increases the collection efficiency and for submicron particles the enhancement is from about 65 to 95 percent. As a result of this, the dependence of particle collection efficiency on particle diameter may disappear with the grade efficiency curve becoming flat.

PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY SHEET NO. _____ OF _____

SUBJECT LEE DESIGN PRESSURE VESSEL PANEL BED

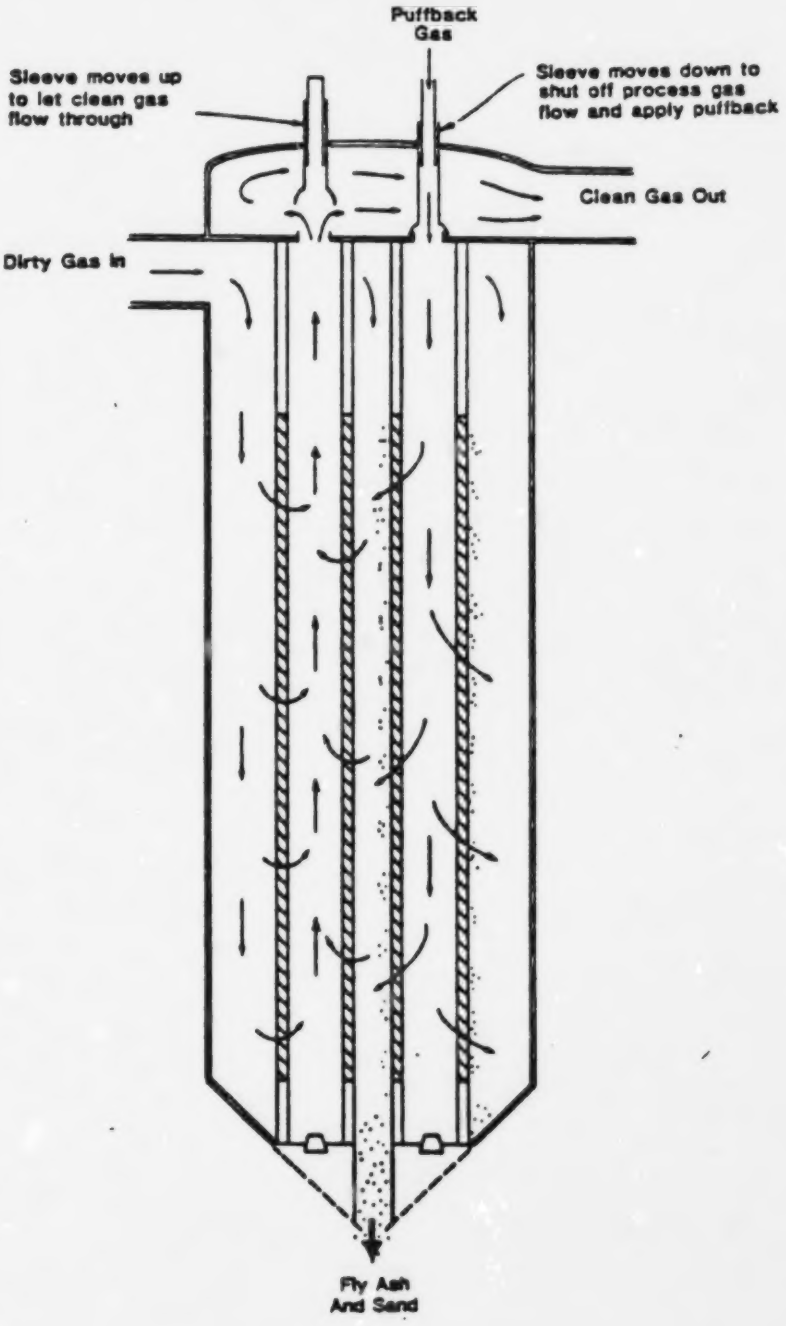
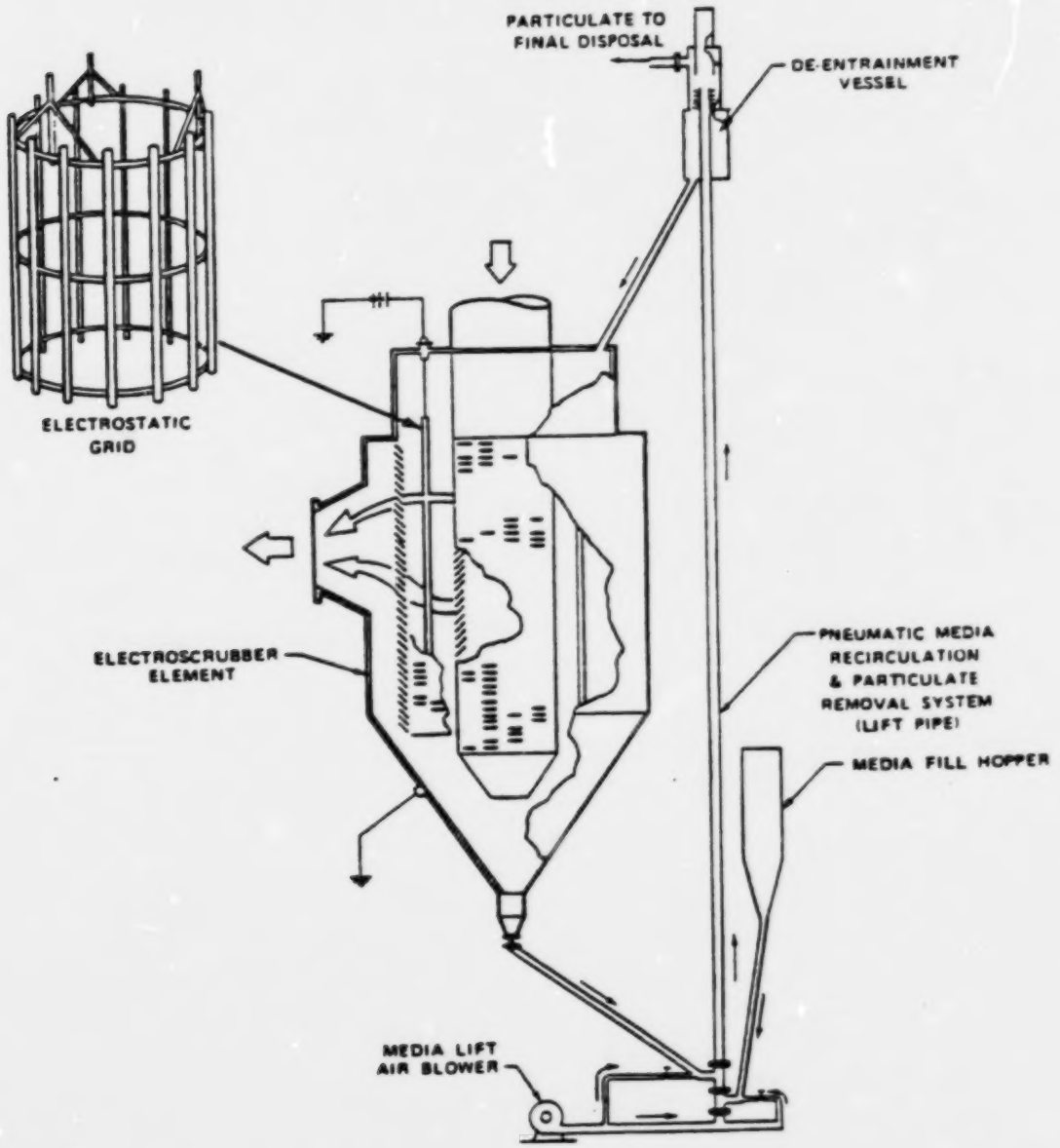

Puffback Gas
Sleeve moves up to let clean gas flow through
Sleeve moves down to shut off process gas flow and apply puffback
Clean Gas Out
Dirty Gas In
Fly Ash And Sand

FIGURE 6.1-7

FIGURE 6.1-8



CROSS-FLOW MOVING-BED ELECTROSCRUBBER GRANULAR FILTER
(COMBUSTION POWER COMPANY, INC.)

6.1.2.5 Novel Technology

Westinghouse is examining the feasibility of hot gas cleaning to minimize the concentration of alkali in a high temperature fuel gas with possible simultaneous particulate removal through the use of alumino silicate materials (getters). Currently the research efforts are focusing on the screening of various alkali-getting materials.

In the screening evaluations, candidate getters were exposed at 850-900°C to approximately 80 ppmv NaCl, in an inert gas atmosphere (argon) for six hours at 1115 kPa pressure. Each of the runs made used an activated bauxite pellet for comparison.

Through the screening evaluations Westinghouse has identified approximately twenty potential candidates for alkali removal.

Further testing by Westinghouse will be acquired to provide adequate information on the performance of the alkali getting materials to be used for combustion turbine protection in a coal gasification combined cycle power generation application.

6.1.3 Evaluation of High Temperature Particulate Removal Technologies

The principal objective of the high temperature particulate removal system is to reduce entrained particulate in the fuel gas (i.e. ash, unburned carbon, alkali, metals) exiting from the three gasification systems to levels which will avoid the plugging of sulphur absorbent vessels and can be acceptable for combustion turbine equipment and environmental emission requirements.

To date high temperature particulate removal technologies are in the research and development stages. The efforts have primarily focused on the removal of particulate entrained in flus gas evolved from a pressurized fluidized bed combustor (PFBC).

The major obstacle to particulate removal analysis for coal gasification is the lack of actual operating data on dust size and loadings in the fuel gas. The particulate matter entrained in flue gas from a PFBC may not be consistent with the particulate size distribution, density and concentrations encountered in a coal gasifier. Therefore, the information provided on collection efficiencies of the various high temperature particulate removal technologies related to the PFBC must be examined with caution.

To date commercially available electrostatic precipitators have operated at atmospheric pressure, although, Research-Cottrell has demonstrated the feasibility of operating at higher pressures. Research-Cottrell has demonstrated with a pilot plant at Curtis-Wright on a pressurized fluidized bed the feasibility of operating at high temperature and pressures. Until further details are received this technology has been set aside for future consideration.

In viewing the preliminary tests obtained for the alternative high temperature particulate removal technologies, the indications are that a combination of cyclones and granular bed filters without electrostatics will not meet the particulate emission standards. This is discussed in Section 6.3. It appears that a combination of cyclones and ceramic bed filters utilized alone may provide adequate protection for the sulphur absorption vessels and combustion turbine equipment, although there may exist the problem of sufficient entrained alkali metal in the fuel gas to cause corrosion and erosion of turbine blading. Therefore, for the purposes of this study a combination of cyclones and ceramic bed filters are recommended for sulphur absorption vessel and combustion turbine equipment protection. In this scheme the cyclones would provide a gross solids separation (>10 microns) with the ceramic fiber filters removing the finer particulate. For alkali control a technology such as that being developed by Westinghouse (See section 6.1.2.1) may be required downstream of the absorbent vessel for combustion turbine protection.

6.2 High Temperature Sulphur Removal Technology

6.2.1 Factors Considered in Technology Evaluation

The principal objective of a high temperature sulphur removal system for coal gasificaion fueled combined cycle power generation is to reduce the sulphur components in the fuel gas to levels which are compatible with environmental emissions and combustion turbine equipment requirements without significant expenditure of energy or loss of efficiency. Ideally, a high temperature sulphur removal system should be capable of removing the sulphur components in the fuel gas to the degree necessary at the temperature leaving the gasifier.

Currently, processes for high temperature sulphur removal are in the research and development stages. Most of the present efforts have employed various metal oxide sorbents in bench scale testing of simulated hydrogen sulphide containing atmospheric pressure, low caloric value (Btu) fuel gas; the primary objective being to meet the sulphur emission standards placed on direct coal fired boilers and, more recently, the operating requirements of molten carbonate fuel cells.

When gasifying coal to produce medium caloric value (MBG) fuel gas, the coal reacts with oxygen and water to produce carbon monoxide, hydrogen and, to a lesser extent, methane and other hydrocarbon gases, with carbon dioxide and unreacted water vapour as the major non-fuel components of the product gas. The feed coal sulphur gasifies primarily to hydrogen sulphide and, to a lesser extent, carbonyl sulphide. A portion of the nitrogen in the feed coal may appear as ammonia and hydrogen cyanide in the fuel gas. The actual composition of the fuel gas evolved is a function of the coal type, gasifier technology and its operating conditions.

Variations in the fuel gas composition may affect the efficiency of a proposed sulphur removal sorbent. These effects can not be accurately predicted and can only be determined by actual operating experience. Therefore, caution

must be exercised when using results obtained from bench scale testing with simulated gases to predict operating conditions which will be encountered in a commercial coal gasification combined cycle power generation application.

In evaluating the various high temperature sulphur removal systems proposed, the following factors should be considered:

6.2.1.1 Absorption Capacity

In use, the metal oxide of the sorbent is converted to the metal sulphide and, while most metal sulphides maintain an ability to decompose organic sulphur compounds, the removal of sulphur as hydrogen sulphide ceases when the oxide is fully converted. At that time, the gas must be diverted to another vessel containing fresh or regenerated sorbent. The maximum capacity of a given sorbent depends on:

- the amount of metal oxide present per unit volume of sorbent which reflects the metal oxide content of the sorbent and its bulk density;
- the availability of the metal oxide in a given sorbent volume which reflects the pore volume of the sorbent and its effective surface area.

These factors are interrelated and may have adverse effects on the performance of a sorbent. For example, while increasing the pore volume or porosity of the sorbent will increase the availability of the oxide, it will also cause a decrease in the bulk density and the mechanical strength of the sorbent. Increasing the pore volume will often also increase the susceptibility for damage to the pore structure by temperature excursions.

For high activity, the sulphur containing gas must enter the interior of the sorbent through the pores. Migration of the sulphide ions into the oxide crystallites as well as the migration of oxide ions out must take place if the oxide present is to be utilized to its fullest. Thus, a definite time is

required for the reactions to occur and the rate of reaction increases with temperature. The overall reaction rate, therefore, is dependent upon the concentrations and types of sulphur compounds, the gaseous space velocity, and reaction temperature.

Ideally, the sulphur would be absorbed by each finite portion of the sorbent bed until it was fully sulphided with a sharp sulphur adsorption front. With only a single, specific sulphur compound present in the gas such ideality could be approached by proper selection of temperature and space velocity. In actuality the sulphur front is diffuse and, while the sorbent at inlet becomes essentially fully sulphided, breakthrough of sulphur usually occurs when the sorbent at the outlet is only partially sulphided. Sulphur breakthrough occurs when the sulphur concentration in the exit gas exceeds a specified level. The absorption capacity of the sorbent is defined as the weighted average sulphur content of the sorbent per unit weight or volume of sorbent at sulphur breakthrough.

6.2.1.2 Regenerability

Any sorbent used for high temperature sulphur removal must be either regenerable or disposable. While most of the potential sorbents, when sulphided, could probably be characterized as non-hazardous, the disposable option has economic disadvantages. These include the cost of the sorbent, the capital investment required to load, unload and transport the sorbent, the labour required for is physical handling and the space required for its inventory and disposal.

If the sorbent selected is not discarded, it must be regenerable. To be considered regenerable, the sorbent and the process or method used for regeneration should satisfy to some extent the following criteria:

- The process should be simple.

- The process should restore the absorption capacity close to that of a fresh sorbent.

- The absorption capacity should not be significantly diminshed with repeated cycles of absorption and regeneration.

- The time required for regeneration should be compatible to that used for the absorption part of the cycle.

- The operting conditions used for regeneration should be commensurate with those of the absorption cycle so as to minimize the need for upgraded materials of construction to suit only the regeneration conditions.

- Regeneraton of the sulphided sorbent should result in a sulphur bearing effluent stream which is readily recovered by commercially available technology.

- The process should be efficient in its use of or need for energy.

- The process should not require the use of agents or chemicals which are not readily and economically available. The use of such agents and chemicals should not impact detrimentally on the environment.

6.2.1.3 Energy Requirements

The incentive for the utilization of a high temperature sulphur removal system for coal derived fuel gas in a combined cycle application is the anticipated increases in cycle thermal efficiencies as compared to the low temperature processes. Therefore, to achieve this objective, a high temperature sulphur removal system should not itself have significant thermal energy losses.

The thermal efficiencies losses in a high temperature sulphur removal system can appear as:

- A direct reduction in the caloric or heating value of the fuel gas.

- The addition of heat input required to maintain some stage of the overall process at an appropriately high temperature.

- Added heat input to maintain an endothermic reaction.

- Losses associated with transportation or exchanging of heat from one reservoir to another.

6.2.1.4 Contaminants

The potential presence of contaminants including entrained particulate matter, tars, ammonia, organic sulphur and nitrogen compounds, alkali compounds, halides and heavy metal compounds can directly affect the absorptive and regenerative properties of a sorbent. Prediction of the presence and identity of such contaminants is difficult and only actual operting experience with the sulphur removal system in association with the coal gasification process will prove the true limitations placed upon the sorbent.

6.2.1.5 Mechanical Strength

A sorbent for commercial application should be sufficiently strong to resist, without appreciable breakdown, various types of stresses, including:

- Resistance to abrasion during transit due to the rolling and tumbling to which it may be subjected to inside a contained drum.

- Resistance to impact loading experienced during the charging of an absorption vessel.

- Sufficient internal cohesion so that gross chemical changes can occur while the sorbent is in use, without these changes causing sorbent break-up.

- Capable of withstanding the load in the sorbent bed due to the weight of the material itself, gas pressure drop, relative movements caused by thermal expansion and contraction of the bed and its containing vessel, and the repeated lifting and settling of the bed material caused by flow reversals.

Attrition resistance is also an important factor in assessing the mechanical strength of a sorbent. Attrition of the sorbent can lead to particle entrainment in the exit gas causing excessive wear of downstream equipment, particularly in a combustion turbine application.

6.2.1.6 Capital and Operating Costs

The capital and operating costs associated with a high temperature sulphur removal technology are a function of the gasifier type, absorptive and regenerative characteristics of the sorbent and the regeneration technique employed. While contributing to the production cost of electric power, the true capital and operaing costs of a developing technology in the conceptual stage are difficult to accurately predict and should not be considered a prime factor in evaluating the alternative technologies. It is of greater importance to consider the capability of the technology to fulfill the overall process specifications.

6.2.1.7 Other Factors to be Considered

In assessing the various high temperature sulphur removal technologies proposed, many other factors can influence the selection of the system, including:

- Materials and equipment requirements;
- Life cycles of the sorbent;
- Availability of the sorbent;
- Adverse side reactions occurring between the fuel gas and sorbent;
- Supporting efforts to develop the system for commercial application; and
- Degree to which the technology has been developed.

6.2.2 Review of Technology

6.2.2.1 Iron Oxide

Process Chemistry

An initial review of the process chemistry and equilibria associated with iron oxide for the removal of hydrogen sulphide in the presence of a reducing gas atmosphere provides an indicator of its potential performance as a sorbent for coal derived fuel gas.

Under the conditions of an oxidizing atmosphere, ferric oxide or hematite (Fe_2O_3) is the stable form of iron oxide. The magnetite form (Fe_3O_4) predominates in a reducing atmosphere and is the sulphur absorbent. The theoretical reactions occurring are:

A. $3\ Fe_2O_3 + H_2 = 2\ Fe_3O_4 + H_2O$

B. $Fe_3O_4 + H_2 + 3\ H_2S = 3\ FeS + 4\ H_2O$

C. $3\ Fe_2O_3 + CO = 2\ Fe_3O_4 + Co_2$

D. $Fe_3O_4 + Co + 3\ H_2S = 3\ FeS + 3\ H_2O + CO_2$

The equilibria between the phases of Fe_2O_3 and Fe_3O_4 are determined by the molar ratios of H_2O:H_2 and CO_2:CO in the reducing gas. Calculated phase diagrams for iron and its oxides at equilibrium in the pesence of H_2, H_2O, CO and CO_2 at various temperatures are illustrated in Figure 6.2-1.

Reactions A and C are cause for concern since they suggest the consumption of hydrogen and carbon monoxide, major components of coal derived fuel gas, during the cycle to reduce the hematite to the magnetite form.

In addition, ferric oxide can be converted to ferrous oxide (FeO) by the reactions:

$$Fe_2O_3 + H_2 = 2\ FeO + H_2O$$

$$Fe_2O_3 + CO = 2\ FeO + CO_2$$

These reactions take place under a reducing atmosphere at temperatures above 625°C dependent upon the molar ratios of H_2O:H_2 and CO_2:CO present in the fuel gas as illustrated in Figure 6.2-1. The concern here is that a ferric to ferrous oxide reduction at the start of the absorption cycle may also affect the quality of fuel gas.

The presence of water vapour in the fuel gas will also affect the equilibrium concentration of hydrogen sulphide in the effluent gas by reactions B and D. The effect of water vapour partial pressure in contact with Fe_3O_4 over the temperature range of 200 to 400°C is illustrated in Table 6.2-1. The basis for the calculations is for a plant operating with a hydrogen:hydrocarbon molar ratio of 0.25 or 20 mole percent hydrogen. As illustrated, the hydrogen sulphide partial pressure is significantly increased by both increases in temperature and the relatively low water vapour partial pressures.

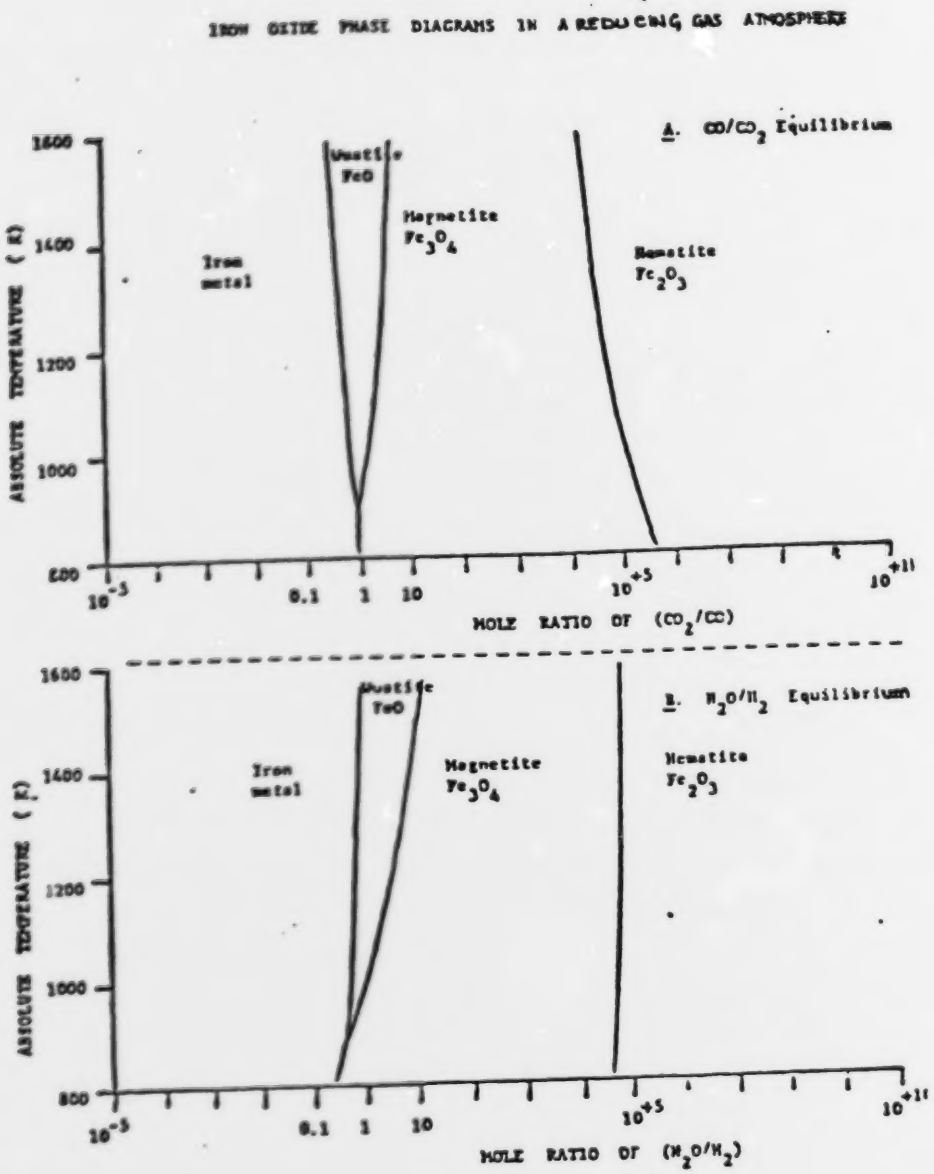
The Morgentown Energy Technology Center (METC) of the U.S. Department of Energy indicated that the absorption of hydrogen sulphide using iron oxide produced both FeS and FeS_2 by the following reaction in addition to reactions B and D:

$$FeS + H_2S = FeS_2 + H_2$$

This reaction liberates hydrogen back into the fuel gas so the overall empirical reaction can be written as:

$$Fe_2O_3 + 3\ H_2S = 2\ FeS_{.5} + 3\ H_2O$$

FIGURE 6.2-1


IRON OXIDE PHASE DIAGRAMS IN A REDUCING GAS ATMOSPHERE
A. CO/CO2 Equilibrium
ABSOLUTE TEMPERATURE (°K)
1600
1400
1200
1000
800
Iron metal
Wustite FeO
Magnetite Fe3O4
Hematite Fe2O3
10-5
0.1
1
10
10+5
10+1
MOLE RATIO OF (CO2/CO)
B. H2O/H2 Equilibrium
ABSOLUTE TEMPERATURE (°K)
1600
1400
1200
1000
800
Iron metal
Wustite FeO
Magnetite Fe3O4
Hematite Fe2O3
10-5
0.1
1
10
10+5
10+1
MOLE RATIO OF (H2O/H2)

Theoretically, the regeneration of iron sulphide to iron oxides can be accomplished in an oxidizing atmosphere to produce either elemental sulphur and/or sulphur dioxide by the reactions:

E. $3\ FeS + 5\ O_2 = Fe_3O_4 + 3\ SO_2$

F. $3\ FeS + 2\ O_2 = Fe_3O_4 + 3\ S$

for ferrous sulphide and similar reactions for pyrite (FeS_2). Excess oxygen converts Fe_3O_4 to Fe_2O_3 by the reaction:

G. $2\ Fe_3O_4 + \frac{1}{2}\ O_2 = 3\ Fe_2O_3$

The regeneration of iron sulphides can also occur in a sulphur dioxide atmosphere to produce elemental sulphur by the reaction:

H. $3\ FeS + 2\ SO_2 = Fe_3O_4 + 5\ S$

J. $3\ FeS_2 + 2\ SO_2 = Fe_3O_4 + 8\ S$

The regeneration of iron sulphide to iron oxide by reactions E, F and G are exothermic, with reactions H and J being endothermic. The use of air or nitrogen-air mixtures as the regenerating media by reactions E and F are cause for concern since they can result in a high temperature rise in the sorbent vessel causing destruction by pore fusing or glassing of iron oxide on its support material.

Experimental Results and Testing

The Morgentown Energy Technology Center (METC), previously the Morgentown Energy Research Center, has been investigating the utilization of solid regenerable sorbents for the removal of hydrogen sulphide from coal derived fuel gas at temperatures above 538°C. The focus of METC's earlier efforts had been

directed towards the evaluation of iron oxide on a range of different support materials to provide a mechanically stable and chemically active sorbent.

The earlier tests conducted at METC, 1974 to 1976, evaluated the use of iron oxide supported on a matrix of coal fly ash with the addition of varying amounts of bentonite to improve physical strength. The coal derived fuel gas was produced utilizing a fixed bed, atmospheric pressure, air blown experimental gasifier. The raw gas was withdrawn from the gasifier and passed through a cyclone separator to remove larger entrained particulates. Then the gas and the hot gas, maintained at approximately 585°C by electric heaters, was contacted with the iron oxide fly ash sorbent in a 15.3 cm diameter, 1.22 m high, stainless steel absorber.

In these tests, four absorption-regeneration cycles were evaluated, based on air blown gasification of a bituminous coal containing 2.71 percent sulphur. The sorbent material tested was a sintered mixture of 25 percent, by weight, iron oxide and 75 percent fly ash with the addition of 3 percent bentonite. Bentonite is a highly swelling clay mineral containing approximately 63 percent SiO_2 and 19 percent Al_2O_3.

During the absorption cycles, the space velocity was maintained at 1910 hourly with a bed inlet temperature of 585°C.

The absorption cycles were terminated when the hydrogen sulphide concentrtion of the absorber effluent gas reached 1.50 grains per SCF, which was considered to be the point of saturation for the sorbent. The concentrations in the effluent gas remained relatively constant at about 0.25 to 0.30 grains of H_2S per SCF of gas for approximately 6 hours and then began to increase at a faster rate until the specified saturation point was reached.

The data indicate 83 percent hydrogen sulphide removal when using the 1.5 gain per SCF point of saturation. By selecting a lower point of saturation, 0.5 grain/SCF, 93 percent hydrogen sulphide removal was achieved. The absorption

capacity for the four test periiods averaged 8.25 weight percent, defined as pounds of hydrogen sulphide absorbed per pound of fresh sorbent. Table 6.2-2 lists the fuel gas analysis and results of the four absorption periods.

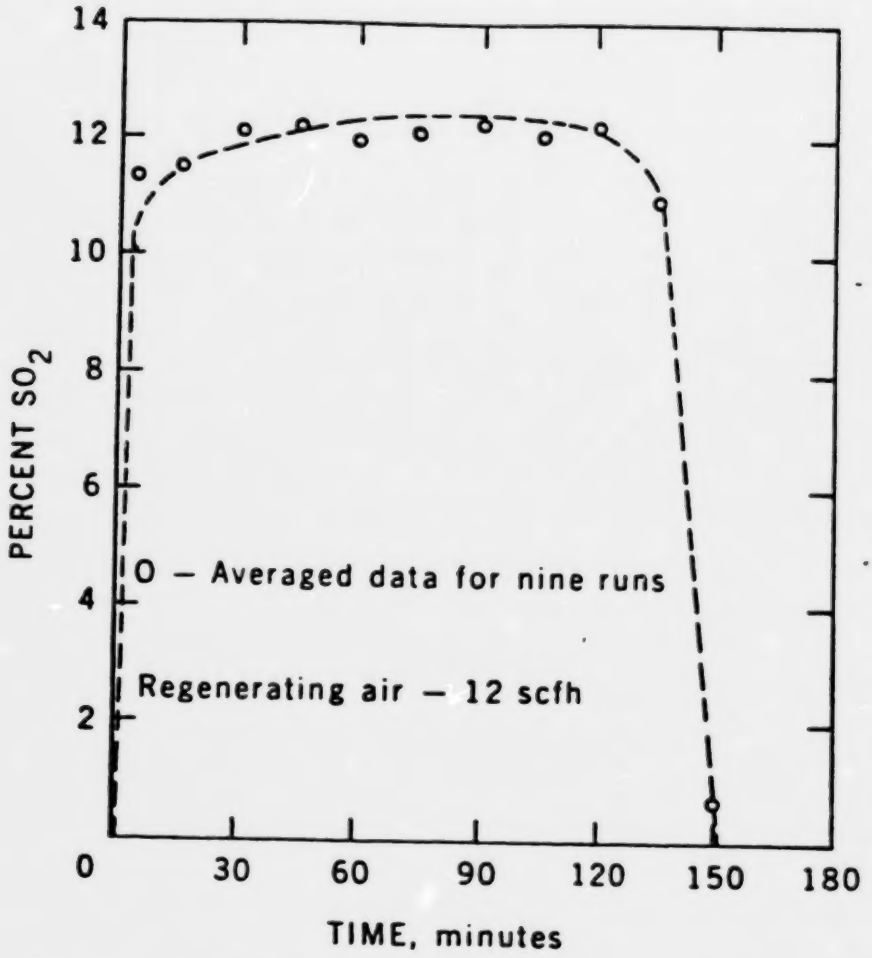
METC encountered problems during regeneration with air, as excess heat evolved could not be dissipated rapidly enough to keep the temperature below 815°C to prevent excessive pore fusing of the iron oxide-fly ash sorbent. Data could not be taken on the sulphur dioxide concentration of the effluent gas but METC indicated that a typical sulphur dioxide concentration curve using an air flow of 12 SCFH can be illustrated by Figure 6.2-2. When using oxygen limited air, an equal mixture of air (12 SCFH) and nitrogen (12 SCFH) for regeneration, the sulphur dioxide concentration would decrease to 6 percent and regeneration time would increase to 240 minutes.

The runs indicated that tars can be tolerated if the absorber bed temperature remains above 540°C. The fine carbonaceous particulate accumulation in the absorber bed, deposited during absorption, is burned off during the regeneration cycle with an insignificant amount of ash residue remaining in the bed after regeneration.

The pressure drop across the bed increased to 60 KPa at the end of the absorption periods but returned to 19 KPa after regeneration of the absorbent bed.

Tests were also conducted to evaluate the chemical and mechanical stability of an iron oxide sorbent supported on fly ash. One sorbent bed went through 30 absorption-regeneration cycles using a simulated raw gas containing coal-related tars and particulate.s The absorbents used in this study were prepared by adding 10 percent by weight of starch to a sorbent mixture containing 25 percent iron oxide and 75 percent fly ash. The raw gas was maintained at 585°C throughout the absorption cycle at an average flow of 15 SCFH. The hydrogen sulphide concentration in the inlet gas stream was maintained at 2 percent and steam content varied between 7 and 11 percent.

FIGURE 6.2-2



Typical curve showing SO_2 concentration in the effluent gas during regeneration of the sulfided sorbents.

Regeneration after each cycle was accomplished by using a 12 SCFH air flow rate for approximately 2 hours. The average absorption cycle lasted 5 hours. After 30 cycles, METC indicated that the sorbent was in good condition without visible signs of deterioration or caking. Table 6.2-3 presents 6 of the 30 runs which were representative of the typical information computed from the raw data obtained during operations.

The influence of various steam concentrations in a coal derived fuel gas were examined by METC to determine its effect on hydrogen sulphide removal efficiency and absorption capacity. The tests concluded that increasing steam concentations would decrease the absorption capacity and efficiency of an iron oxide fly ash mixture as a sulphur sorbent. The tests also showed that conditions existed during the abosrption cycle for the water gas shift reaction to occur with the iron oxide acting as a catalyst.

The water gas shift reaction occurs as:

$$CO + H_2O = CO_2 + H_2$$

Table 6.2-4 shows the results of four absorption tests at two temperatures and four steam concentrations and compares the calculated hydrogen content of the outlet gases shifted to equilibrium with the measured hydrogen content. As illustrated, the tests at 595°C and 705°C with 13.4 and 11.7 percent steam concentrations, respectively, demonstrate the change in gas composition affected by the shift reaction. At 595°C, the carbon monoxide and water content is reduced by about 3.45 percent, while hydrogen and carbon dioxide content are increased correspondingly. At 705°C, the change is about 1.16 percent. Comparison of the calculated and measured hydrogen content of the effluent stream shows that the water gas shift reaction is essentially complete.

The tests at 595°C and 705°C with 15.4 and 20.8 percent steam concentrations, respectively, show a decrease in absorption capacity to the increased steam concentrations. At 595°C, the carbon monoxide and water content is reduced by

about 4 percent with hydrogen and carbon dioxide increasing accordingly. The corresponding change in composition at 705°C is about 1.75 percent. The calculated and measured hydrogen content of the effluent stream also show that the shift reaction approaches the equilibrium concentration.

Research efforts were continued to develop a support and binding material which would combine high crushing strength, high sulphur pick-up and still permit the higher temperature operaton required for regeneration. In these efforts, varying amounts of silica, bentonite and sodium silicate were sintered and tested for crush strength, abrasion resistance and sulphur loading capacity. It was concluded that a 45 percent, by weight, iron oxide and 55 percent silica using sodium silicate as a binding agent appeared to have higher absorption capacity, crushing strength and an apparent higher heat resistance than the fly ash supported sorbents. It was suggested by METC that the iron oxide silica supported sorbent could be operated up to temperatures of 930°C during regeneration without damaging the sorbent and would have an average absorption capacity of 12 to 14 percent by weight. A summary of the absorption capacity and mechanical strength tests for fly ash and silica supported iron oxide sorbents is presented in Tables 6.2-5 and 6.2-6, respectively.

Beginning in 1980, a series of experimental tests were conducted by METC to determine the capability of zinc oxide-based sorbents to remove sulphur compounds from hot coal derived fuel gas. The program was an extension of the investigations based on the development of the iron oxide sorbents. The principal objective of this investigation was the development of a regenerable sorbent capable of reducing the sulphur content of the hot fuel gas to less than about 1 ppm, a requirement for coal gas to be used as a fuel for molten carbonate fuel cells.

Before proceeding to test the zinc oxide based sorbents, tests were first performed with an iron oxide-silica sorbent developed by METC in conjunction with Air Products and Chemicals to evaluate the regenerative capabilities of

various steam air mixtures. Air Products and Chemicals had experimented with mixtures of air and steam for sorbent regeneration and had shown that, in addition to moderating regeneration temperatures, steam mixed with air in certain proportions converts part of the sulphur from the spent iron oxide sorbent into elemental sulphur vapour in the regeneration of gases.

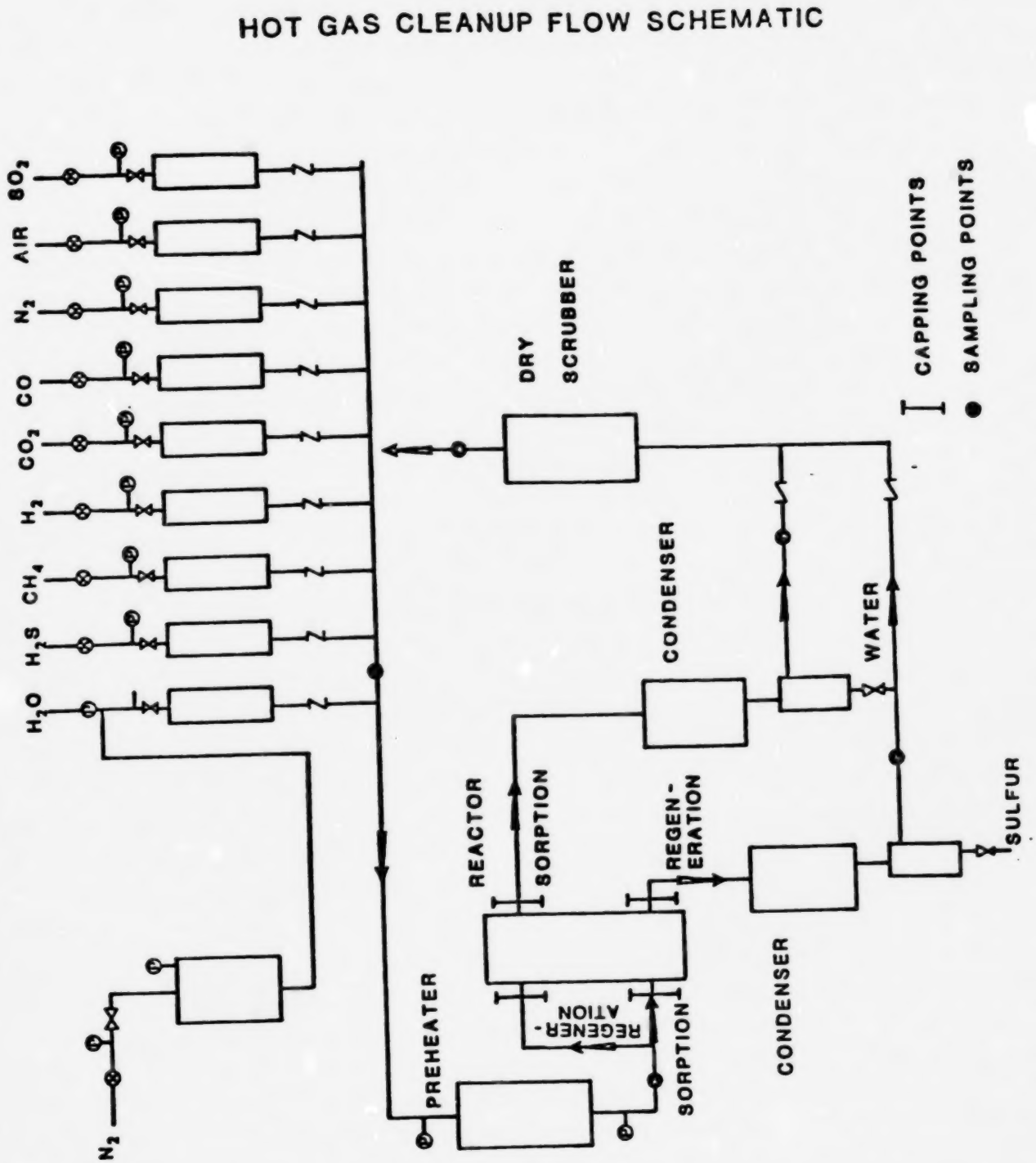
Central to the test unit was a 5.1 cm diameter by 0.8 m long reactor made of stainless steel. The reactor was equipped with shell type electric heaters. As installed, its operating pressure was limited to about one atmosphere above ambient, with operating temperatures above 980°C. A schematic diagram of the test unit is illustrated in Figure 6.2-3.

Cylinders of various gases required to simulate a coal derived fuel gas were used in the testing. Steam was added to the gas mixture by means of a pressurized tank of distilled water. The reactor effluent gases were passed into a water cooled condenser where the steam was condensed. In the regeneration mode, air and steam pass through the reactor preheater where they are heated to the operating temperatures.

Since the tests were mainly concerned with the regenerative and absorptive capabilities of the various metal oxides rather than mechanical strength, the sorbents were prepared by adding small amounts of bentonite, water and extruding. The iron oxide based sorbent contained 45 percent, by weight, iron oxide with the remainder being mostly silica.

The tests were completed in three cycles of sulphidation and regeneration with three separate loadings of the iron oxide sorbent at the reactor inlet temperatures of 538°C. For sulphidation, the gas mixture used was a simulated METC fixed bed gasifier coal gas similar to that previously employed, except that carbon oxides and hydrocarbons were replaced by nitrogen. For sulphidation, a space velocity of approximately 800 to 1000 hourly was used throughout the testing. Sulphidation was continued until sulphur loadings were about 16 to 19 percent of fresh sorbent. For regeneration, three different mixtures of

FIGURE 6.2-3


HOT GAS CLEANUP FLOW SCHEMATIC
AIR
SO_2
N_2
CO
CO_2
H_2
CH_4
H_2S
H_2O
N_2
DRY SCRUBBER
CAPPING POINTS
SAMPLING POINTS
CONDENSER
WATER
REACTOR
SORPTION
REGEN-ERATION
SULFUR
PREHEATER
REGENER-ATION
SORPTION
CONDENSER

steam and air were examined; namely, 95/5, 90/10 and 50/50 percent by volume. The space velocity for regeneration was 600 hourly.

Table 6.2-7 summarizes the sulphidation-regeneration operating conditions and results of the tests conducted with the iron oxide silica based sorbent as reported by METC.

In Test 1, after sulphidation, a mineral analysis was performed, revealing that the iron content of the sulphidated sorbent was in the form of pyrite, hematite and magnetite, roughly in the ration of 1:1:2.

Test 2 was a regeneration with 95/5 percent steam-air mixture heated to 538°C. The run was continued for 39 hours during which virtually all the absorbed sulphur was liberated with about 75 percent of it appearing as elemental sulphur. A sulphur analysis of the regenerated sorbent confirmed that regeneration was virtually complete.

For the second regeneration, Test 4, the regenerant was a 90/10 percent mixture of steam-air. Elemental sulphur was about 15 percent of the recovered total, the remainder being a mixture of hydrogen sulphide and sulphur dioxide in comparable proportions.

A third regeneration was carried out in Test 6 with a 50/50 percent steam-air mixture. In this case, elemental sulphur represented only about 1.5 percent of the total sulphur with the bulk of the sulphur being liberated as sulphur dioxide accompanied by a small amount of hydrogen sulphide.

There was little noticable difference in surface area, density and pore volume between sulphidated and regenerated sorbent. They were, respectively, about 1.0 m^2/gm, 3.25 gm/m^3 and 2.2 mm^3/g.

A surprising result was the low initial exit levels of hydrogen sulphide before breakthrough in the sulphidation runs. This level was an order of magnitude less than that previously obtained and comparable to that achieved in later tests with zinc oxide based sorbents. It was concluded that, since the iron oxide tested was a by-product of the steel industry, the presence of zinc oxide determined to be approximaely 7 percent by weight of the sorbent, contributed to the low hydrogen sulphide concentrations.

In spite of substantial elemental sulphur formation when regenerating with steam-air mixtures lean in air, METC concluded that economics suggest that processes based on this for sulphur recovery would probably not be feasible in most cases. This conclusion reflects the large steam requirements and the potential difficulty of condensing sulphur and disposing of the effluent steam. Since the addition of 50 percent steam in air produced a satisfactory moderating effect on regeneration temperatures, METC decided to conduct future regeneration tests with zinc oxide based sorbents using the mixture as the regenerant.

6.2.2.2 Zinc Oxide

Zinc oxide sorbents for the desulphurization of process streams have been commercially used for many years. The major applications have been in the final desulphurization of hydrocarbon feedstocks to steam-hydrocarbon reforming units, and as sulphur guard vessels for the protection of copper-zinc low temperature water gas shift catalysts. Bulk sulphur removal is usually achieved upstream by liquid absorption processes. When utilized in the above applications, the sulphided zinc oxide is normally discarded after sulphur absorption capacity is achieved.

The reaction occurring between zinc oxide and hydrogen sulphide is:

$$ZnO + H_2S = ZnS + H_2O$$

The effect of water vapour partial pressure on the equilibrium hydrogen sulphide partial pressure in contact with zinc oxide over the temperature range of 450 to 750°C is illustrated in Table 6.2-8.

It is known that zinc oxide is difficult to reduce by the reactions:

$$ZnO + H_2 = Zn + H_2O$$

$$ZnO + CO = Zn + CO_2$$

in a normal reducing gas atmosphere, suggesting that the consumption of the major coal derived fuel gas components will not occur, although if zinc metal formation occurs at temperatures near 725°C, vapourization would proceed immediately.

Generally, in order to avoid cracking of the heavier hydrocarbons, a zinc oxide sorbent (typically containing 80 weight percent zinc oxide) is not subjected to temperatures higher than 400°C. Imperial Chemical Industries (ICI) has reported that a catalyst operated at 370°C and a space velocity of 400 hourly until breakthrough normally achieves an average sulphur absorption capacity in the range of 22 to 24 percent, by weight, of sulphur, equivalent to 24.4 to 26.7 kg of sulphur per 100 kg fresh catalyst. From a design point, under the above operating conditions, ICI suggests a sulphur content coresponding to an average of 18 percent by weight (equivalet to 20 kg sulphur per 100 kg charged) be considered as a practical level. In practice, when utilized as a low temperature shift catalyst guard, a zinc oxide sorbent is operated closer to 250°C and the average sulphur absorption capacity before breakthrough is significantly lower.

METC evaluated the performance of a commercial, nominal 100 percent zinc oxide sorbent with the following characteristics:

Apparent Bulk Density	1.2 g/cm^3
Crushing Strength	3.6 kg
Surface Area	3 m^2/g
Pore Volume	0.26 cm^3/g

A summary of the operating conditions and results for the zinc oxide sorbent tested are presented in Table 6.2-9. The testing procedures and equipment used are the same as those described in Section 6.2.2.1 for the iron oxide silica sorbent as related to Figure 6.2-3.

Tests 1 to 3, two sulphidations and one regeneration at 538°C, demonstrated that the zinc oxide sorbent, after one regeneration, lost none of its absorption efficiency or capacity for the test conditions employed. Under the conditions used, the sorbent reduced the hydrogen sulphide content of the gas from 2.7 percent down to less than 10 ppm.

Regeneration with a 50/50 percent mixture of steam-air produced an off gas with 3 to 14 percent sulphur dioxide and 0.16 grams of elemental sulphur on a dry basis.

The surface area, pore volume and density were measured before Run 1, fresh, and after Run 3, sulphidized. Results showed a decline of surface area and pore volume from 3.2 m^2/g and 6.5 m^3/g to 1.1 m^2/g and 2.0 mm^3/g, respectively. The density changed from 6.4 g/cm^3 to 3.8 g/cm^3. A mineral analysis of the sulphidized sorbent indicated a composition of 15 percent ZnO and 85 percent ZnS on a weight basis.

Tests 4 to 12 were conducted to determine the effect of temperature on the absorptive power of zinc oxide. The absorption and regeneration space velocities employed were 2000 and 600 hourly, respectively.

In Run 4 at a temperatue of 427°C, the hydrogen sulphide concentration was brought down to 3 ppm for 1.5 hours. Upon regeneration in Run 5, Runs 6 and 8 at 538°C and 649°C, gave similar hydrogen sulphide removal efficiencies but a higher absorption capacity, approximately 7.5 weight percent, although the absorption capacity was less than half that achieved with a space velocity of 1000 hourly. A noticeable departure in absorption capacity became evident in Run 10 at 760°C, where the hydrogen sulphide removal level was virtually the same but capacity was less than half as much at about 2 percent. At this point, regeneration Test 11 was aborted, as sulphur dioxide was being evolved at much lower concentrations. Sulphidization Test 12 at 816°C with a new batch of sorbent was similar to Run 10 in both the hydrogen sulphide levels in the absorption off-gas and absorption capacity.

The sulphur dioxide concentrations ranged from 3.0 to 14.8 percent on a dry basis for all the regeneration runs with the 50/50 volume percent steam-air mixture.

6.2.2.3 Zinc Ferrite

In light of the knowledge gained during METC's development program for a regenerable iron oxide sorbent, zinc ferrite surfaced as a candidate to reduce hydrogen sulphide concentrations of coal derived fuel gas to the levels consistent with molten carbonate fuel cell limitations. This compound has the same crystal structure as the magnetite form of iron oxide (i.e. the spinel structure) and is synthetically produced by heating together iron oxide and zinc oxide powders in the correct molar proportions. It was reasoned that this material could have a combination of the absorption efficiency of zinc oxide and the capacity and regenerability of iron oxide. It is also known from metallurgical practice that zinc ferite is difficult to reduce and it was concluded that there would be a lesser probability of elemental zinc volatization than would be expected in the case of zinc oxide.

METC evaluated the absorptive and regenerative capabilities of zinc ferrite in comparison to the capabilities of iron oxide and zinc oxide based sorbents. A summary of the salient conditions and results of the experiments are presented in Table 6.2-10. The experimental procedures and equipment utilized in the testing are equivalent to that discussed for iron oxide in Section 6.2.2.1 as related to Figure 6.2-3.

Tests 1 to 6, with the same charge of sorbent, had as their purpose determination of the hydrogen sulphide removal capability of zinc ferrite and verification of its regenerability. As in the tests with iron oxide, both sulphidation and regeneration were at 538°C. The regenerant was a 50/50 percent by volume mixture of steam and air.

For all three of the sulphidation runs, the hydrogen sulphide was removed to a level of less than 6 ppm and, after two regenerations, there was no noticable loss of absorption capacity.

During regeneration, a reaction zone moved through the sorbent bed for about 8 hours accompanied by a temperature peak of up to 954°C for Test 2, with the regeneration off gas containing about 5 to 10 percent sulphur dioxide by volume on a dry basis. It was reported that significant amounts of hydrogen sulphide, in the order of 1000 ppm, were also present, as well as small quantities of elemental sulphur.

Surface area, pore volume and density determinations were made before Test 1, after Test 2 and after Test 5B. Over the series there was a fall off in surface area from 5.8 m^2/g (initial unsulphidized) to 1.4 m^2/g (final, sulphidized). Correspondingly, pore volume declined from 12.2 mm^3/g to 2.9 mm^3/g and density changed from 5.2 g/cm^3 to 4.4 g/cm^3.

The sorbent was analyzed for sulphur after Run 1 and after Run 5B. It was determined that the sulphur content averaged 0.4 and 26.4 percent by weight, respectively. A mineral analysis was conducted after Run 5B and it was deter-

mined that the sorbent averaged 28.5 percent FeS, 38.5 percent ZnS, and 33.0 percent $ZnFe_2O_4$. In contrast with a corresponding analysis for the iron oxide sorbent, no pyrite (FeS_2) was detected.

Tsts 7 to 22 were conducted to determine the absorption performance of zinc ferrite over a range of temperatures from 316°C to 816°C. All regenerations were at a nominal temperature of 538°C.

For Tests 7 and 8 at 538°C, both absorption and regeneration space velocities were doubled to 2000 and 1200 hourly, respectively, in order for METC to complete tests to breakthrough within an 8-hour shift. METC reported that the higher space velocity for regeneration proved to be injudicious after a satisfactory absorption run in 7, caused by a temperature peak of over 1038°C. In Run 9, at 316°C with a new loading of sorbent, hydrogen sulphide breakthrough occurred immediately. After approximately 2 hours, the inlet gas temperature was raised to 538°C and hydrogen sulphide was reduced to 100 ppm. At this point the sorbent was regenerated and a further absorption run attempted at 538°C. Initially, hydrogen sulphide came through a 7 ppm but breakthrough occurred within half an hour.

It is reasoned from Run 9 that the kinetics of absorption at 316°C are poor to reach equilibrium hydrogen sulphide levels, although the hydrogen sulphide flow rate in the simulated gas may be an order of magnitude higher than would be expected in practice.

The reactor was filled with a new loading of zinc ferrite in Test 10. In Tests 10 and 11 absorption and regeneration were performed at 538°C, with a regeneration space velocity at 600 hourly and both tests produced equivalent results previously obtained. Test 12, a repeated sulphidation at 538°C, again proved normal. In future runs, the lower regeneration space velocity of 600 hourly was re-adopted.

After regeneration in Test 13, Test 14 was conducted at 427°C, with hydrogen sulphide exiting the reaction at 5 ppm initially and at 900 ppm after half an hour. It is reasoned that reaction kinetics are better than 316°C but still poor at this temperature. Because of the short duration of absorption, the subsequent regeneration was included in the same test. In Test 15, the absorption temperature was raised to 649°C and the hydrogen sulphide concentration was down 9 to 15 ppm for 2.5 hours. After the fourth regeneration, there was an increase in temperature to 760°C in Run 17. Hydrogen sulphide removal before breakthrough was superior to that at 649°C and comparable to that at 538°C but absorption capacity was lower with breakthrough occurring after 1.5 hours.

The next sulphidation, after regeneration in Test 18, was at 816°C in Run 19 when hydrogen sulphide was initially removed to the 4 ppm level but for only half an hour. Test 20 was performed again at 816°C with a fresh batch of sorbent because, before Test 19, the sorbent had been inadvertently overheated to 871°C. The results of this test are, however, much the same.

To test whether the sorbent was damaged at 816°C, a further sulphidation in Test 22 was conducted after regeneration in Test 21. It was observed that breakthrough had occurred immediately. The surface area and pore volume for Run 22 were 2.2 m^2/g and 5.2 mm^3/g, suggesting that sintering may not have been the main reason for poor performance.

A mineral analysis after Run 19 gave 48.5 percent FeO, 28.7 percent ZnS, 5.3 percent $ZnFe_2O_4$ and 17.4 percent ZnO, indicating that most of the ferirc iron in zinc ferrite had been reduced to ferrous oxide. There was no sulphur absorbed as FeS, contrary to that experienced at lower temperatures. Another mineral analysis, after Test 22, gave 91.8 percent $ZnFe_2O_4$ and 8.2 percent ZnS. It is reasoned that the sorbent had been reconverted to zinc ferrite after sulphidation at 816°C and subsequent regeneration, although its absorption capacity was relatively small, indicating irreparable through unaccountable damage.

6.2.2.4 Other Processes

Theoretically, calcium-based compounds can remove hydrogen sulphide from a coal derived fuel gas dependent upon the hydrogen sulphide and carbon dioxide content and temperature of the gas stream. Thermodynamically, both calcium oxide and calcium carbonate are capable of reacting with hydrogen sulphide to produce a stable sulphide, CaS. No other sulphides are possible with calcium. The absorption would proceed by the following reactions which are dependent on the water vapour and carbon dioxide content:

$$CaO + H_2S = CaS + H_2O$$

$$CaCO_3 + H_2S = CaS + H_2O + CO_2$$

The carbon dioxide partial pressure of the coal-derived fuel gas defines the temperature at which limestone calcines, as given by the phase boundary in Figure 6.2-4.

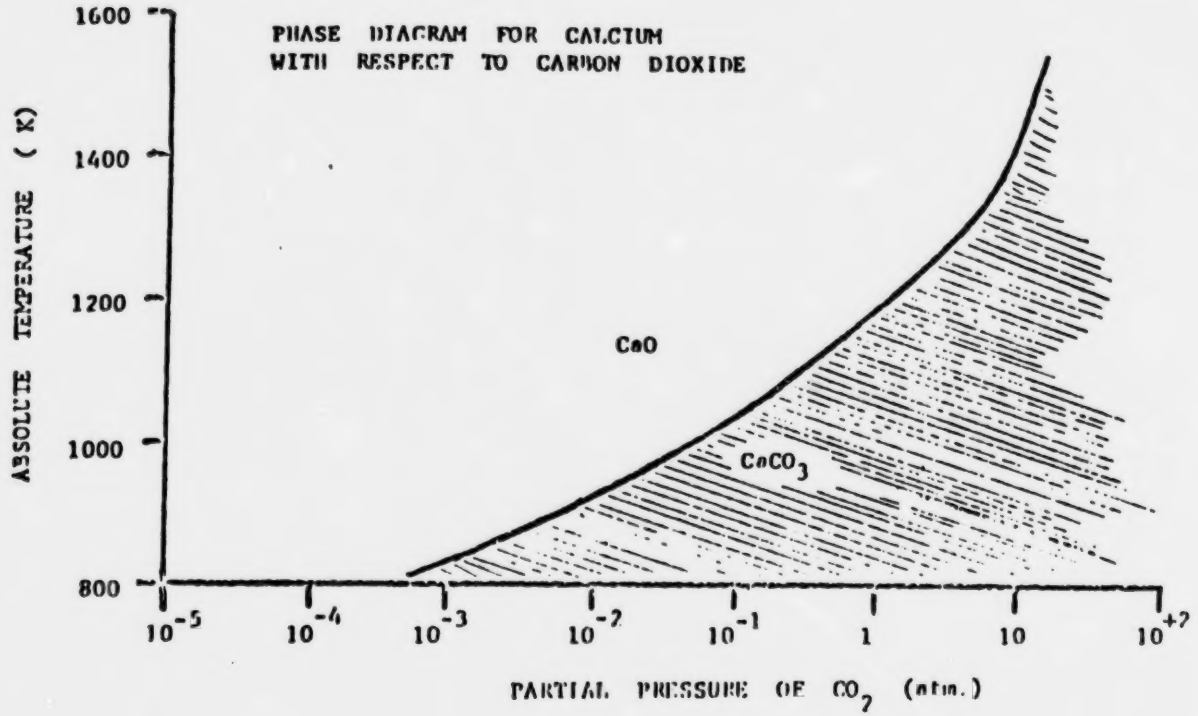
It has been reported by METC in a review of the various metal oxide sorbents available, that, kinetically, the reactivity of calcium-based sorbents behave differently from thermodynamic predictions and that $CaCO_3$ is kinetically inert to H_2S while CaO reacts rapidly to form CaS. Dolomite reacts with H_2S in its half-calcined state by:

$$CaMg(CO_3)_2 = CaMg(CO_3) + CO_2$$

$$CaMg(CO_3) + H_2S = CaMgS(CO_3) + H_2O$$

When fully calcined, dolomite is not as reactive with H_2S as is either calcined limestone or half-calcined dolomite. With dolomites, the hydrogen sulphide absorption temperature is near 850°C.

FIGURE 6.2-4



6-40A

Since $MgCO_3$ frequently coexists with $CaCO_3$, its chemistry was considered. As with calcium, neither free Mg metal nor $Mg(OH)_2$ can be formed in a coal derived fuel gas atmosphere. $MgCO_3$ is calcined very rapidly, so that only MgO should exist at the conditions of a hot coal-derived fuel gas. The reaction with hydrogen sulphide could proceed as:

$$MgO + H_2S = MgS + H_2O$$

$$MgO \cdot CaCO_3 + H_2S = MgO \cdot CaS + CO_2 + H_2O$$

Upon further evaluation of the thermodynamics and kinetics associated with calcium and magnesium-based sorbents, it was concluded by METC that neither system appeared promising as an effective absorbent for hydrogen sulphide removal from coal derived fuel gas.

Recently, METC has awarded contracts for the bench scale testing and development of a lithium aluminate sorbent and a mixed iron oxide process to remove hydrogen sulphide and hydrogen chloride from coal-derived fuel for use in molten carbonate fuel cells. The processes would operate at approximately 650°C.

In the mixed bed iron oxide process, the metal oxides to be investigated include zinc and iron orthosilicate, iron chromite and cobalt aluminate. It is claimed that, upon regeneration, elemental sulphur can be produced directly by heating the oxides to between 815 and 980°C.

In the lithium aluminate process, a fixed bed of pellets made of lithium aluminate could be filled with a molten mixture of mixed salts. The salts are expects to be lithium, potassium or sodium carbonate or a mixture of the three. Preliminary work has suggested that the salts would be capable of removing both hydrogen chloride and sulphur to the 1 ppm levels.

An alternate concept to the absorption of sulphur-bearing gases evolved upon the gasification of coal, is the insitu desulphurization of coal within the gasification process. Rockwell International, over the past ten years, has been developing a Molten Carbonate Coal Gasification (Rockgas) Process to produce a sulphur-free fuel gas.

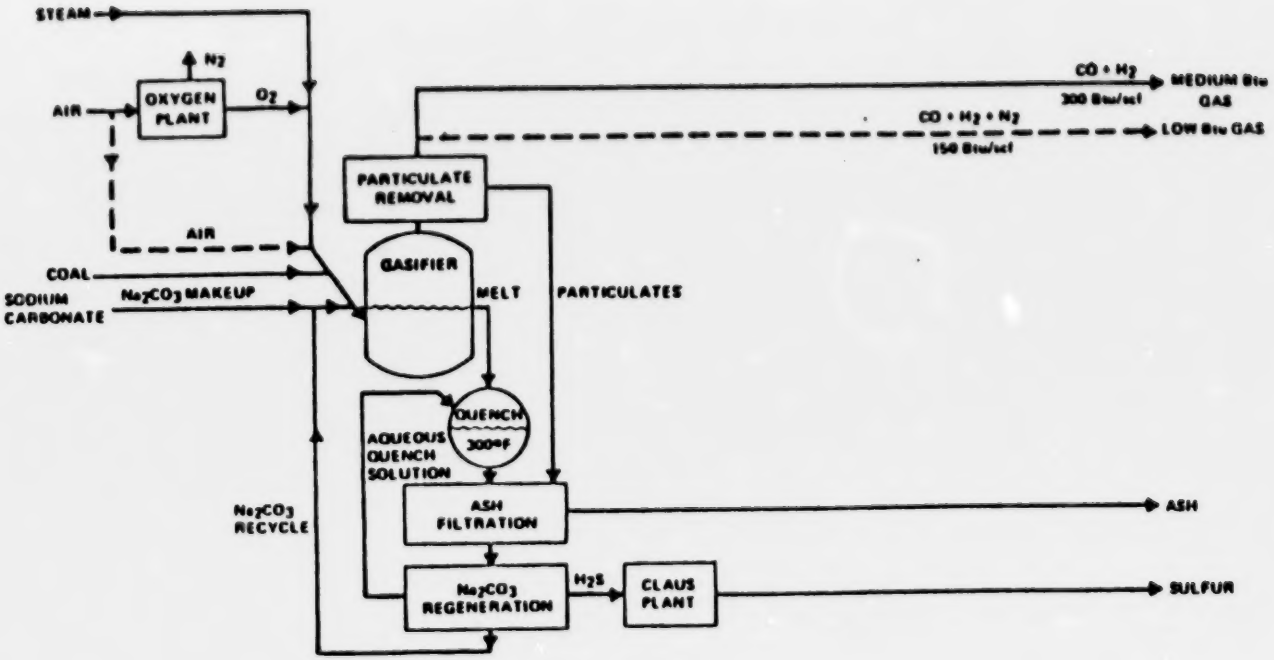
A generalized flow diagram of the Rockgas process is shown in Figure 6.2-5. The diagram illustrates the production of low- or medium-Btu gas by using either air or high purity oxygen as the oxidant. Typical calculated gas compositions for the two cases are given in Table 6.2-11.

In the process, coal is gasified in a turbulent pool of sodium carbonate at a temperature of approximately 930°C and pressure in the range of 5 to 20 atmospheres. The coal is injected beneath the surface of the melt pool together with the required sodium carbonate makeup. The sulphur reacts with sodium carbonate to form sodium sulphide, which is claimed to have a catalytic effect on the gasification reactions.

A small stream of melt is continuously removed from the gasifier and replaced by the addition of dry sodium carbonate with the coal. The melt removed from the gasifier is immediately quenched by contact with a water solution. Sodium carbonate and sodium sulphide dissolve in the hot solution, while unreacted carbon and insoluble ash constituents remain as undissolved particles.

The slurry is then processed to remove the insoluble ash components, strip off sulphur as hydrogen sulphide, and precipitate sodium bicarbonate crystals. The ash is disposed of as a wet filter cake, the hydrogen sulphide can be converted to elemental sulphur in a conventional Claus plant, and the sodium bicarbonate crystals are calcined to convert them to dry sodium carbonate powder suitable for recycling.

FIGURE 6.2-5



Rockwell International

Simplified Flow Diagram of Molten Salt Coal Gasification Process

Rockwell has operated a process development unit capable of gasifying approximately 22 metric tons per day of coal and has successfully demonstrated the technical feasibility of the process to produce an essentially sulphur-free low Btu fuel gas at approximately 930°C and the regenerability of the sodium carbonate/sulphide melt.

The major obstacle to the direct use of the hot desulphurized fuel gas from the Rockgas process in a combined cycle application is the possibility of sodium carbonate/sulphate entrainment in the fuel gas, thereby potentially causing excessive corrosion and erosion of combustion turbine blading.

A detailed assessment of the applicability of the Rockgas process is beyond the scope of this report, but should be considered as a viable alternate concept for further study.

6.2.3 Evaluation of High Temperature Sulphur Removal Technologies

To determine whether high temperature gas purification will increase the overall thermal efficiency of combined cycle power generation as compared to the currently proposed low temperature gas purification processes, a reference high and low temperature process must be selected. For the purposes of this study, it was concluded (Section 5) that a physical solvent process such as the Selexol Process appears to be technically feasible for comparison evaluations.

In reviewing the preliminary results obtained by METC, both zinc and iron oxide based sorbents are capable of reducing the hydrogen sulphide concentrations produced in the three different coal gasifiers as required by the product gas specifications established in Section 4.3. In addition, both zinc and iron oxide based sorbents are regenerable using controlled concentrations of oxygen.

The absorption capacity of an iron oxide sorbent is adversely affected by the presence of water vapour as determined by METC in their testing (see Section 6.2.2.1). Therefore, iron oxide may impose a penalty on Lurgi gasification by suppressing hydrogen sulphide removal efficiencies. The multivalent nature of iron might, it has been postulated by METC, cause consumption of hydrogen and carbon monoxide by reducing the hematite form of iron oxide to magnetite. Zinz oxide based sorents would not be expected to have such a problem unless the operating conditions would reduce zinc oxide to significant amounts of metallic zinc.

Zinc oxide based sorbents have been shown to be capable of reducing the hydrogen sulphide levels in a fuel gas to approximately 5 to 10 ppm at space velocities for which iron oxide would result in 200 to 800 ppm.

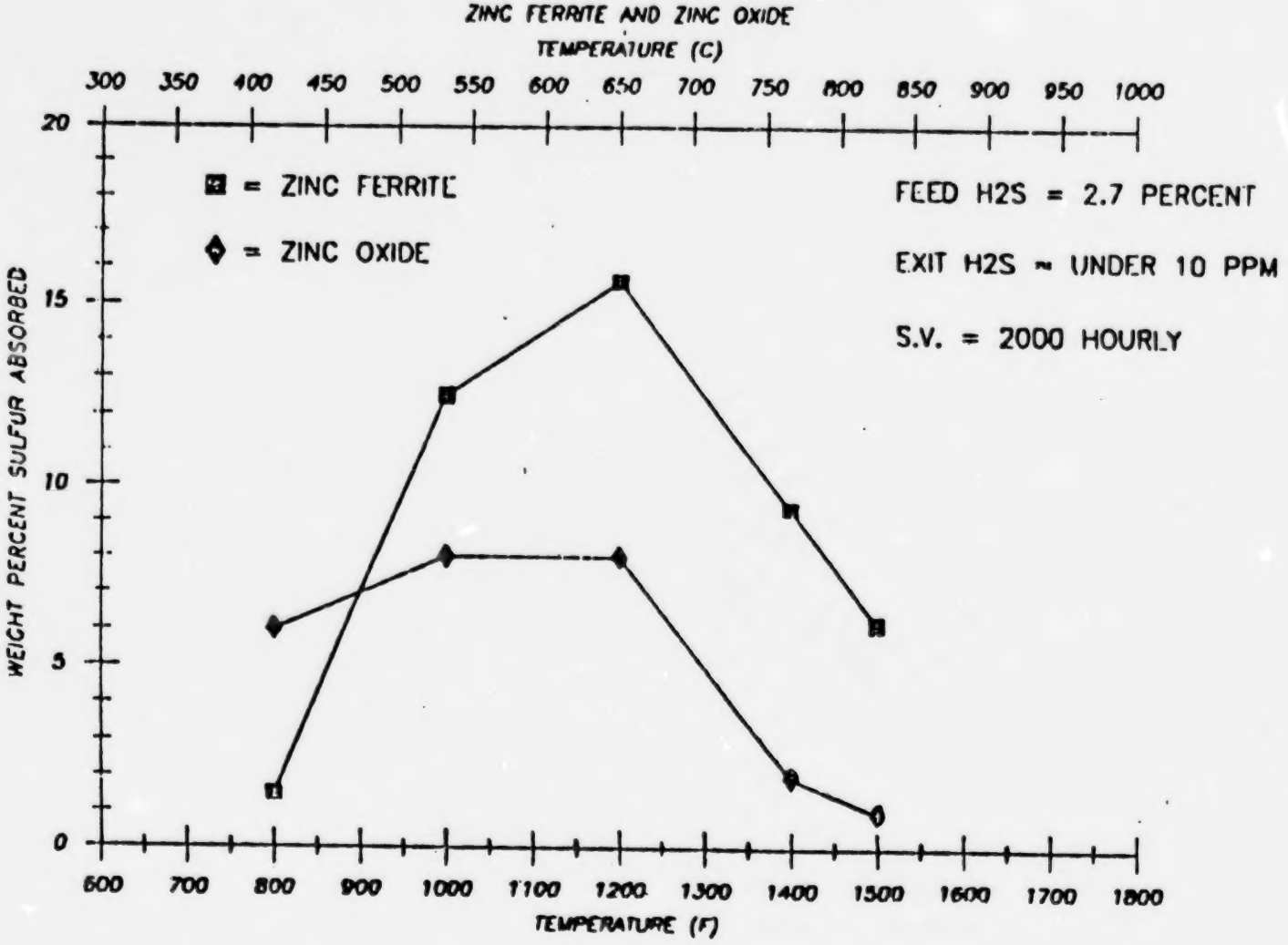
In comparing zinc oxide and zinc ferrite sorbents, METC concluded that zinc ferrite appears to have higher absorption capacities at equivalent space velocities in the higher temperature range as illustrated in Figure 6.2-6. Zinc ferrite also is reported to be more resistant to sintering than zinc oxide and iron oxide.

As discussed in Section 6.2.2, the regeneration of the metal oxides will produce an effluent stream with a low concentration of sulphur dioxide, although it has been demonstrated that steam lean in air can produce elemental sulphur. To avoid the expense of using steam to produce elemental sulphur or moderate the regeneration temperature rise, which can cause destruction of the sorbent. Conceptually, a recycle sulphur dioxide air mixture could produce a gas suitable for sulphuric acid production.

Of the studied metal oxides, zinc ferrite sorbents appear to be the most favourable for further development and testing for possible commercial application. For the purposes of this study, zinc ferrite will be selected as the representative high temperature sulphur removal technology. The method of sorbent regeneration will incorporate the use of a sulphur dioxide mixture

FIGURE 6.2-6

SULFUR LOADING


ZINC FERRITE AND ZINC OXIDE
TEMPERATURE (C)
300 350 400 450 500 550 600 650 700 750 800 850 900 950 1000
20
15
10
5
0
WEIGHT PERCENT SULFUR ABSORBED
= ZINC FERRITE
= ZINC OXIDE
FEED H2S = 2.7 PERCENT
EXIT H2S ~ UNDER 10 PPM
S.V. = 2000 HOURLY
600 700 800 900 1000 1100 1200 1300 1400 1500 1600 1700 1800
TEMPERATURE (F)

with the recovery of the resulting sulphur dioxide by sulphuric acid production.

This choice of the representative high temperature sulphur removal process should not be considered as the optimum process for all gasification technologies and applications. The general conclusion is that any process development should cover a sensitivity study on the interaction of the gas composition and sorbent performance for the full range of gas compositions expected from each gasifier to which application is proposed.

6.3 High Temperature 'Clean-Up' Technology for Coal Gasification

The composition of gas exiting the three types of coal gasifier covers a wide range of gaseous products and particulate, due to the gasifier characteristics. This can be attributed to the differing coal feed size, mixing behaviours, temperature, pressure and throughput that influence the design of a "clean-up" technology for the gasification process. The high temperature gas clean-up removes the contaminants from the gas without any significant cooling.

The clean-up criteria for the impurities depends largely on meeting the environmental standards and satisfying particular process requirements. Gas turbine manufacturers specifications are generally more stringent than environmental standards. Contaminants levels in the coal gas fuels must be closely maintained in order that particulates do not exceed six microns in size with the average particle size not exceeding three microns. Trace sulphur compounds in the fuel gas form sodium and potassium sulphates in the gas turbine and cause corrosion-erosion and deposition, and affect blade life over 649°C (1200°F).

Although there is insufficient information available relating to the particulate emissions standards for coal gasification combined cycle power plants, it was realized that the particulate emission standards for the British Columbia Hydro PFBC Combined Cycle Project or the United States Environmental Protection Agencies New Source Performance Standards (NSPS) for coal fired boilers can be adopted for the purposes of this study. The limits for particulate emissions are 0.07 lbs/MM.Btu for B.C. Hydro and 0.03 lbs/MM.Btu for the NSPS. Ebastec's approach has incorporated the EPA NSPS because this is approximately the low particulate emission level from Ontario Hydro nignite or coal fired plants. It is recommended for high temperature

fuel gas purification that a combination of cyclones and ceramic fibre filters for particulate removal and zinc ferrite for sulphur dioxide removal be installed. It appears that a combination of cyclones and granular bed filters without electrostatics will not meet either of the particulate emission approaches. Supporting data is provided in reference 2 of the bibliography, "Dust Removal Concepts for Pressurized Fluidized Bed Systems", by R.R. Boericke and W.E. Moore. This reviews the alternative high temperature particulate removal systems. The performance data (Figure 18) indicate that it is impractical to design a granular bed filter to meet environmental emissions at 0.07 lbs/MM.Btu without resorting to electrostatics. Economic data in this reference presented in Table V indicate a relatively narrow cost advantage (4 percent) for using cyclones plus electrostatics ganular bed filters vesus cyclones plus ceramic filters. Since the particulate removal system is 4 to 7 percent of the total plant investments, the capital cost advantage would translate into less than 1 percent of the total plant investment for the High Temperature Gas Purification cases. Therefore, based on this information, it appears that the selection of ceramic fiber filters should not unduly penalize the High Temperature Gas Purification schemes proposed.

Although high-temperature gas clean-up systems have the advantage of potentially increasing the thermal operating efficiency, a considerable amount of development work is required before the required level of harmful contaminant removal can be achieved. Most of these components for cleaning coal gas at high temperature are not in a form that can be guaranteed for commercial operation. It is difficult to ascertain what operating life and reliability they have attained.

Westinghouse Electric Corporation in the EPRI funded "High-Reliability Gas Turbine Combined-Cycle Development" study have stated the following limits on coal fuel gas contaminants:

Fuel Gas Contaminants to the Combustor (PPMV)							Fuel Gas Particulate
	H_2S	HCl	KCl	NaCl	NH_3	COS	(Grains/Scf) (Case A)
Hot-Gas Clean-up	700*	150	3.4**	1.5**	320		.018

* Assumes 90% sulphur removal by the Hot-Gas Clean-Up System

** Does not take into account any reduction due to getters

SECTION 6 - CONCLUSIONS AND RECOMMENDATIONS

(a) The recommended process scheme for high temperature fuel gas purification for the three gasification technologies will include a combination of cyclones and ceramic fiber filters for particulate removal and zinc ferrite as the sulphur removal sorbent. The particulate removal system would be oriented upstream of the sulphur absorption vessels to avoid potential plugging. Dependent upon the anticipated performance of the particulate removal system and attrition of the zinc ferrite sorbent a polishing filter may be required upstream of the combustion turbine equipment.

(b) For the purposes of this study it will be assumed that a technology such as that being developed by Westinghouse utilizing an alumino silicate getter would provide adequate protection for the combustion turbines. In the capital and operating cost estimates it is only possible to make an allowance for this technology which has yet to be developed.

TABLE 6.2-1

EFFECT OF WATER VAPOUR AND TEMPERATURE ON EQUILIBRIUM OF HYDROGEN SULPHIDE IN CONTACT WITH IRON OXIDE[1]

H_2O (%)	Equilibrium H_2S (ppm v/v)				
	Temperature (°C)				
	200	250	300	370	400
3.3	1.85	3.1	5.1	8.9	10.8
1.7	7×10^{-1}	1.35	2.2	3.51	4.23
0.33	0.85×10^{-2}	1.57×10^{-1}	2.01×10^{-1}	4.01×10^{-1}	4.9×10^{-1}
0.17	1.57×10^{-3}	0.62×10^{-1}	0.93×10^{-1}	1.64×10^{-1}	1.96×10^{-1}

Note: 1. Reaction

$$Fe_3O_4 + 3\ H_2S + H_2 = 3\ FeS + 4\ H_2O$$

Gas contains 20 mole percent hydrogen

TABLE 6.2-2

RAW GAS ANALYSIS FOR FOUR ABSORPTION PERIODS

Run	Percent, Dry Basis					
	CO	CO_2	H_2	N_2	CH_4	H_2S
1	17.3	13.0	17.8	46.8	3.6	0.48
2	16.2	12.2	15.7	52.5	3.6	0.58
3	23.2	6.7	13.5	52.5	3.2	0.52
4	20.9	8.1	13.5	56.1	3.1	0.53

RUN DATA OBTAINED DURING FOUR ABSORPTION PERIODS

	Run 1	Run 2	Run 3	Run 4
Volume of bed (cu. ft.)	0.785	0.785	0.785	0.785
Weight of absorbent (lbs.)	73	73	73	73
Gas flow rate (scfh)	1,500	1,500	1,500	1,500
Space velocity scfh/cu. ft. absorbent)	1,910	1,910	1,910	1,910
Duration of run (hrs.)	15.87	9.78	14.50	10.75
Inlet steam concentration (%)	8.5	3.3	3.1	6.3
Bed temperature (°C)	585	585	585	585
H_2S absorbed (grains)	46,950	31,532	53,624	47,450
H_2S absorbed when 1.50 grains = point of saturation (%)	83	83	81	86
H_2S absorbed when 0.50 grains = point of saturation (%)	92.9	91.4	90.7	94.2
Absorption capacity (weight-percent)	9.19	6.17	8.35	9.29

Conversion Factor: 1 SCF = 29,215 Ncm^3

TABLE 6.2-3

TYPICAL DATA FROM SIX OF THIRTY ABSORPTION CAPACITY TESTS USING FLY ASH-IRON OXIDE SORBENT

Gas Flow Rate (Std. cu. ft./hr.)	Run Duration (hr.)	Space Velocity (vol./vol./hr.)	H_2S Absorption			Total Quantities Through Bed	
			Total Grains	Weight (%)	Efficiency (%)	Tar (gm)	Dust (gm)
14.27	6.28	501	1118	10.35	92.5	31.5	.2
15.97	5.75	563	1193	11.04	95.0	30.8	.6
16.37	5.75	570	1180	10.92	93.5	28.2	.4
16.38	5.50	587	1135	10.51	93.1	33.5	.5
15.23	6.00	529	1159	10.73	94.2	32.3	.5
15.09	7.25	549	1332	12.33	92.2	33.4	.5

Absorption temperature - 595°F
Saturation breakthrough point - 400 gr/100 scf
Weight of absorbent - 700 grams

TABLE 6.2-4

EFFECT OF TEMPERATURE AND STEAM CONCENTRATION ON H_2S ABSORPTION CAPACITY

	Temp. °C	Steam (%)	H_2S Absorbed		Effluent Gas Composition (%)[3]					Hydrogen[4] Concentration Effluent Gas
			Capacity[1]	Efficiency[2]	CO	N_2	CO_2	H_2O	H_2	
Input	-	13.41	-	-	13.4	44.0	6.64	13.41	22.59	-
Output	595		13.43	93.5	9.9	44.0	10.1	9.95	26.04	24.90
Input	-	11.7	-	-	13.66	44.8	6.77	11.69	23.0	-
Output	705		13.03	96.08	12.50	44.8	7.93	10.53	24.16	24.16
Input	0	15.4	-	-	13.13	43.13	6.51	15.4	22.14	-
Output	595		12.82	95.05	9.11	42.83	10.52	11.39	26.15	24.00
Input	-	20.83	-	-	12.24	40.21	6.06	20.82	20.64	-
Output	705		9.36	93.54	8.77	40.24	9.53	17.53	24.11	22.55

[1] Capacity expressed as grams H_2S absorbed per 100 grams sorbent.

[2] Efficiency calculated as percent absorbed of total H_2S through bed.

[3] Calculated from input gas composition assumin complete removal of H_2S and shifting to equilibrium.

[4] From infra-red analyzer

All inputs refer to gas composition entering the absorber.

All outputs refer to gas composition leaving the absorber.

TABLE 6.2-5

EFFECT OF ADDITIVES ON SORPTION CAPACITY AND PHYSICAL PROPERTIES OF FLY ASH-IRON OXIDE SORBENTS

Sorbent Composition	Sintering Conditions					
75% Fly Ash, 25% Iron Oxide with Binder Shown Below	Temp. (°C)	Time (min.)	Surface Area (m^2/gm)	Hardness Retained on 4-mesh (%)	Crush Strength (lb./cm)	H_2S Sorption Capacity Weight (%)
None	980	30	.900	-	16.4 ± 6.7	11.15
Bentonite, 0.5%	980	30	1.073	93.6	53.9 ± 13.9	11.90
Bentonite, 1.0%	980	30	1.156	94.6	53.2 ± 14.2	11.70
Bentonite, 1.5%	980	30	1.503	93.0	59.4 ± 7.3	10.50
Bentonite, 3.0%	980	30	1.349	90.0	88.6 ± 6.9	6.50
Bentonite, 1.0% + 10% starch	980	30	-	-	44.5 ± 10.6	19.70
Bentonite, 1.0% + 20% starch	980	120	1.780	-	46.2 ± 12.5	12.20
Sodium silicate, 10 ml/lb	980	30	-	94.9	69.9 ± 35.8	10.40
Sodium silicate, 10 ml/lb + 10% starch	980	30	-	-	26.7 ± 6.1	14.58
Sodium silicate, 10 ml/lb + 10% starch	980	120	-	-	-	10.93
Magnesium sulphate, 2.0%	980	30	-	-	34.4 ± 14.5	8.10
Magnesium sulphate, 2.0% + 10% starch	980	30	-	-	31.6 ± 16.8	7.70

TABLE 6.2-6

SORPTION CAPACITY AND PHYSICAL PROPERTIES
OF VARIOUS GRADE 1160 SILICA-IRON OXIDE SORBENTS

Sorbent Number	Description of Sorbent	Sintering Conditions				
		Temp. (°C)	Time (min.)	Crush Strength (lb./cm)	Surface Area (m^2/gm)	H_2S Capacity gm H_2S/100 gm Sorbent
1	75% silica + 25% Fe_2O_3	980	30	13.7 ± 4.9	-	8.9
2	65% silica + 35% Fe_2O_3	980	30	19.3 ± 7.6	-	12.5
3	55% silica + 45% Fe_2O_3	980	30	33.8 ± 6.1	4.3	15.4
4	55% silica + 45% Fe_2O_3	1010	30	39.6 ± 19.2	3.2	16.6
5	45% silica + 45% Fe_2O_3	1035	30	77.4 ± 15.8	2.9	15.1
6	55% silica + 45% Fe_2O_3	1065	30	75.5 ± 13.9	2.5	16.1
7	55% silica + 45% Fe_2O_3, 0.5% bentonite	980	30	57.4 ± 11.8	3.8	15.1
8	55% silica + 45% Fe_2O_3, 1.0% bentonite	980	30	31.4 ± 6.7	4.3	14.5
9	55% silica + 45% Fe_2O_3, 10 ml S.S./lb.	980	30	48.2 ± 7.1	3.1	22.4
10	55% silica + 45% Fe_2O_3, 6.25 ml S.S./lb.	980	30	68.0 ± 24.0	2.1	22.8
11	55% silica + 45% Fe_2O_3, 7.5% starch + 20 ml S.S./lb.	980	30	55.3 ± 23.7	2.2	14.3
12	55% silica + 45% Fe_2O_3, 10% starch + 10 ml S.S./lb.	980	30	11.9 ± 2.6	2.4	17.3
13	55% silica + 45% Fe_2O_3, 10% starch + 10 ml S.S./lb.	980	120	54.9 ± 22.3	3.3	16.3
14	65% silica + 35% Fe_2O_3, 10% starch + 10 ml S.S./lb.	980	30	96.4 ± 27.9	2.6	11.17
15	55% silica + 45% Fe_2O_3 (Fisher)	980	30	26.1 ± 6.4	4.1	14.4
16	55% silica + 45% Fe_2O_3 (MCB)	980	30	33.8 ± 6.1	4.6	15.4
17	55% silica + 45% Fe_2O_3 (Fisher), 10% starch	980	30	11.4 ± 2.9	3.8	17.1
18	55% silica + 45% Fe_2O_3 (MCB), 10% starch	980	30	15.3 ± 3.8	3.8	16.6

S.S. = sodium silicate

TABLE 6.2-7

SUMMARY FOR EXPERIMENTAL RESULTS AND CONDITIONS FOR IRON OXIDE

Test Number	Test Type [4]	Sorbent Wt Gm	Sulphur Loading Before Break-through gm sulphur/gm fresh sorbent	Gas Feed Composition Vol%	Space Velocity Vol/ Vol-Hr.	Sorption Time Before Break-through or Regeneration time (hr)	Exit Gas Composition
1	Sulphidation	936		Note 1	800	8	85 ppm H_2S
2	Regeneration	936		95% Steam 5% Air	600	39	7.2% H_2S No SO_2 122 gm elemental sulphur
3	Sulphidation	936	0.0447	Note 2	1013	4.5	22 ppm H_2S
4	Regeneration	936		90% Steam 10% Air	622	20.5	5000-4000 ppm H_2S; 0.3% ppm SO_2; 19.4 gm sulphur
5	Sulphidation	1000	0.104	Note 2	900	3	1 to 6 ppm H_2S
6	Regeneration	1000		50% Steam 50% Air	600	7.5	1000-1200 ppm H_2S 4-7% SO_2 2.5 gm sulphur

Note:
1. 0.7% H_2S, 7% H_2O, 15% H_2, 77.3% N_2
2. 2.7% H_2S, 6.9 % H_2O, 14.6% H_2, 75.8% N_2
3. Exit Gas Composition Dry Basis; SO_2 Initial to maximum
4. All tests were at 538°C and atmospheric pressure

TABLE 6.2-8

Volume % Steam in Gas	H_2S Concentration (ppm) in Wet Exit Gas					
	450°C	500°C	500°C	600°C	650°C	700°C
50	1.7	3.7	7.3	13.2	22.0	34.5
40	1.4	3.0	5.9	10.6	17.6	27.6
30	1.0	2.2	4.4	7.9	13.2	20.7
20	0.69	1.5	2.9	5.3	8.8	13.8
10	0.35	0.75	1.5	2.6	4.4	3.5
5	0.17	0.37	0.7	1.3	2.2	3.5

TABLE 6.2-9

SUMMARY OF EXPERIMENTAL RESULTS AND CONDITIONS FOR ZINC OXIDE

Test Number	Test Type[4]	Sorbent Weight (gm)	Sulphur Loading gms Sulphur per gms Fresh Sorbent before Breakthrough	Gas Composition	Space Velocity (vol./vol./hr.)	Temp. (°C)	Sorption Time Before Breakthrough or Regeneration Time (hrs.)	Exit Gas[3] Composition
1	Sulphidation	1000	0.187	Note 1	1000	538	9	5-9 ppm H_2S
2	Regeneration	1000	0.188	50% air 50% steam	600	538	12	3-14% SO_2 0.16 gm elemental sulphur
3	Sulphidation	1000		Note 1	1000	538	9	2-7 ppm H_2S
4	Sulphidation	1000		Note 2	2000	427	1	3 pm H_2S
5	Regeneration	1000		50% steam 50% air	600	538	6	5-14% SO_2
6	Sulphidation	1000		Note 2	2000	538	2	1-3 ppm H_2S
7	Regeneration	1000		50% steam	600	538	7.5	3.7-14.8% SO_2
8	Sulphidation	1000		Note 2	2000	649	0.25	6-15 pm H_2S
9	Regeneration	1000		50% steam 50% air	600	538	6.75	3.7-14.8% SO_2S
10	Sulphidation	1000		Note 2	2000	760	0.5	2-3 pm
11	Regeneration	1000		50% steam 50% air	600	538	19	regeneration aborted
12	Sulphidation	1000		Note 1	2000	816	0.25	2 ppm H_2S

NOTES:
1. 2.7% H_2S, 6.9% H_2O, 14.6% H_2, 75.8% N_2
2. 2.7% H_2S, 9.78% CO_2, 27.66% H_2O, 11.68% CO, 28.17% N_2, 17.11% H_2, 2.9% CH_4
3. Exit Gas Composition Dry Basis SO_2; Initial to Maximum
4. All Tests at Atmospheric Pressure

TABLE 6.2-10

SUMMARY OF EXPERIMENTAL RESULTS AND CONDITIONS FOR ZINC FERRITE

Test Number	Test Type[4]	Sorbent Weight (gm.)	Sulphur Loading gms. Sulphur per gms. Fresh Sorbent before Breakthrough	Gas Composition	Space Velocity (vol./vol./hr.)	Temp. (°C)	Sorption Time Before Break-through or Regeneration Time (hrs.)	Exit Gas[3] Composition
1	Sulphidation	800	0.213	Note 1	1000	538	7	1-2 ppm H_2S
2	Regeneration	800		50% air 50% steam	600	538	12.5	2000 ppm H_2S 4-9% SO_2
3	Sulphidation	800	0.187	Note 1	1000	538	6	2-4 ppm H_2S
4	Regeneration	800		50% steam 50% air	600	538	12.5	2000-3000 ppm H_2S 4-10% SO_2
5A	Sulphidation	800		Note 1	1000	538	0.5	
5B	Sulphidation	800	0.214	Note 1	1000	538	7	2-6 ppm H_2S
6	Regeneration	800		50% steam 50% air	600	538	12.5	2000-3000 ppm H_2S 4-10% SO_2
7	Sulphidation	643	0.141	Note 2	2000	538	2.25	9-13 ppm H_2S
8	Regeneration	643		50% steam 50% air	1200	538	5	8-13% SO_2 7.7 gm elemental sulphur
9	Sulphidation	643		Note 2	2000	316		6000 ppm H_2S immediately
10	Sulphidation	643		Note 2	2000	538	2.5	3-10 ppm H_2S
11	Sulphidation	643		50% steam 50% air	600	538	6.75	3.5-14% SO_2 4.1 gm elemental sulphur

TABLE 6.2-10 (Cont'd)

SUMMARY OF EXPERIMENTAL RESULTS AND CONDITIONS FOR ZINC FERRITE

Test Number	Test Type[4]	Sorbent Weight (gm.)	Sulphur Loading gms. Sulphur per gms. Fresh Sorbent before Breakthrough	Gas Composition	Space Velocity (vol./vol./hr.)	Temp. (°C)	Sorption Time Before Break-through or Regeneration Time (hrs.)	Exit Gas[3] Composition
12	Sulphidation	643	0.125	Note 1	2000	538	0.2	3-5 ppm H_2S
13	Regeneration	643		50% steam 50% air	600	538	6.75	3.5-13.6% SO_2
14	Sulphidation	643	0.0156	Note 2	2000	427	0.5	5 ppm H_2S
15	Sulphidation	643	0.156	Note 2	2000	649	2.5	9-13 ppm H_2S
16	Regeneration	643		50% steam 50% air	600	538	5.25	3.1-12.8% SO_2
17	Sulphidation	643	0.094	Note 2	2000	760	1.5	1-3 ppm H_2S
18	Regeneration	643		50% steam 50% air	600	538	6.5	2.9-12.8% SO_2
19	Sulphidation	643	0.0626	Note 2	2000	816	0.5	4 ppm H_2S
20	Sulphidation	643		Note 2	2000	816	0.25	2 ppm H_2S
21	Regeneration	643		50% steam 50% air	600	538	1.25	4-5% SO_2
22	Sulphidation	643		Note 2	2000	538	0.25	600 ppm H_2S

NOTES:
1. 27% H_2S, 6.7% H_2O, 14.6% H_2, 75.8% N_2
2. 27% H_2S, 27.66% H_2O, 9.78% CO_2, 11.68% CO, 2.9% CH_4, 28.17% N_2
3. Exit Gas Composition Dry Basis; SO_2 Initial to Maximum
4. All Tests at Atmospheric Pressure

TABLE 6.2-11

TYPICAL GAS COMPOSITIONS FOR LOW- AND MEDIUM-BTU GAS PRODUCTION BY MOLTEN SALT GASIFICATION OF ILLINOIS NO. 6 COAL

	Gasification	
	Air	Oxygen-Steam
Gas Composition (volume %)		
CO	28.7	57.2
H_2	12.0	32.6
CH_4	1.7	2.7
CO_2	2.8	3.7
H_2O	1.8	3.3
N_2	53.0	0.5
H_2S (ppm)*	110	260
Gas Heating Value (Btu/scf)	148	317

* at 10 atm

7. DESCRIPTION OF REFERENCE PLANTS

Development of technology to satisfactorily purify coal-derived fuel gas at elevated temperatures would potentially increase the efficiency of combined cycle power generation over that feasible with current low temperature purification means.

Raw gas produced by the gasification of coal for use in combined cycle power plants typically leaves the gasifier at 520 to 1320°C and over 2000 kPa pressure. The gas is contaminated with particulate matter, nitrogen compounds, and sulphur compounds which must be reduced in concentration, before firing in a combustion turbine, in order to protect the environment and the gas turbine.

Current technology for purifying the gas required cooling the gas typically to temperatures below 150°C in order to remove the suphur compounds by scrubbing with a suitable solvent. Coincidentally, effective removal of particulates is achieved with the cooling and scrubbing. Having the gas cool also permits removal of a source of NOX emissions by scrubbing it to remove ammonia.

The need to cool the gas before purification makes effective use of the heat removed difficult. Only some of the heat can be recovered as steam for use at steam cycle efficiencies of less than about 35 percent. Logically, purification of the raw gas at temperatures nearer those at the exit of the gasifiers would eliminate much of the thermal and power losses involved in cooling the gases.

Thus the investigation and development of hot gas purification is motivated by the premise that the thermal losses associated with it will be considerably lower than those associated with cooled gas purification. Offsetting such gains, however, are the thermal and power losses involved with the

regeneration of the solid metal oxide sorbent used for the high temperature sulphur removal.

Three different gasification technologies are considered to indicate how the incentive to adopt high temperature purification may be influenced by the mode of gasification. The Lurgi (dry bottom) technology represents gasification in a movng bed at "low" temperatures. Typifying gasification in a fluidized bed at "moderate" temperature is the Westinghouse technology while Texaco technology is used to represent entrained flow gasification at "high" temperature.

Each of the three gasifiers have been considered in association with low and high temperature fuel gas purification. The Selexol solvent process (Norton Company) is used as representative of low temperature sulphur removal. For high temperature sulphur removal a conceptual process, based on data developed by METC, is used which utilizes zinc ferrite as the solid regenerable sulphur solvent.

The technique for cooling the crude gas and removal of particulate matter therefrom is normally provided as part of the gasification technology package. This was done in this Study for each of the low temperature purification cases. Particulate removal in all the high temperature purification cases utilizes a suitable combination of cyclones and ceramic fiber filters.

It should be recognized that the purpose of this Study is not to compare different gasification technologies, different low temperature purification processes, or different high temperature purification means. Attention is focused on comparing specific processes for high and low temperature purification in conjunction with a given gasifier operating at a given set of conditions. Comparisons between gasification technologies should not be attempted. Such comparisons would require a detailed study of each gasifier technology and would focus on comparing optimized modes of operation for each gasifier in relation to each type of purification.

To shorten and simplify the references to the various cases considered in this Study, the following letter code utilizing the first two letters is used:

Case ML - Moving bed gasification, Low temperature purification. This case uses the Lurgi dry ash gasifier and the Selexol process for sulphur removal.

Case MH - Moving bed gasification, High temperature purification. This case uses the Lurgi dry ash gasifier and solid zinc ferrite for sulphur removal.

Case FL - Fluidized bed gasification, Low temperature purification. This case uses Westinghouse gasification process with Selexol for sulphur removal.

Case FH - Fluidized bed gasification, High temperature purification. This case uses Westinghouse gasification with zinc ferrite for sulphur removal.

Case EL - Entrained flow gasification, Low temperature purification. This case uses Texaco gasification process with Selexol for sulphur removal.

Case EH - Entrained bed gasification, High temperature purification. Uses Texaco gasification with zinc ferrite for sulphur removal.

7.1 Process Descriptions

For each case studied the following documents were prepared and are presented together with the corresponding process descriptions:

- Simplified process schematic for the plant
- Simplified steam system schematic for the plant
- Material balance
- Overall energy balance

These documents are all on the basis of gasifying a 100 tonnes of coal per hour and consequently show differing power productions.

7.1.1 Lurgi Dry Bottom Gasifier with Low Temperature Purification

(Case ML)

A grass roots facility for producing medium Btu fuel gas by gasifying 100 metric tonnes per hour of Illinois No. 6 coal employing the oxygen-blown Lurgi Dry-Ash Gasification Process is presented on the process flow schematic Figure 7.1-1. Sulphur removal is achieved through the low temperature Selexol process and the facility is integrated with a Combined Cycle Plant for the ultimate production of electric power.

The facility consists of the following sections:

- o Air Separation
- o Lurgi Dry-Ash Gasification Process
- o Gas Cooling
- o Selexol Sulphur Removal
- o Sulphur Recovery Plant
- o Combined Cycle Plant

Necessary offiste and environmental treatment units required for the facility's operation are included in the design.

Reference should be made to Table 7.1-1 for the Material Balance of this case and to Table 7.1-1 for the Overall Energy Balance.

7.1.1.1 Air Separation

The oxidant feed system provides 98 mole percent oxygen to the gasification process at 161°C and an absolute pressure of 2450 kPa. Atmospheric air is compressed, cooled and subsequently processed in a cryogenic air separation unit into 98% purity oxygen. The oxidant is then recompressed to the required pressure in a three stage centrifugal compressor with interstage cooling. Approximately 47.5 metric tonnes per hour of oxidant is produced and fed to the Lurgi gasifiers.

7.1.1.2 Lurgi Dry-Ash Gasification Process

The Lurgi gasifier is a fixed bed, oxygen-blown cylindrical vessel mounted on a vertical axis. For this design the gasifier operates at approximately 2440 kPa pressure and 581°C. The reactor shell is enclosed by a water jacket which serves to control the wall temperature and to recover heat by producing medium pressure steam. This steam generated in the jacket provides a small portion of the steam required for the gasification reaction itself.

Sized coal (38 mm x 6.4 mm) is conveyed from a coal storage area to the gasifier and deposited into the coal hopper/lockhopper system located on top of the vessel. The lockhopper is pressurized (with tail gas, mainly N_2) and the coal is transferred to the reactor through a hydraulically operated valve. The coal moves slowly down the reaction vessel through several distinct reaction zones countercurrent to the flow of gas. In the first zone at the top of the gasifier, coal is preheated and dried by contact with the hot crude gas exiting the reactor. As the coal moves down and is heated further, devolatilization occurs and gasification commences. The bottom of the bed is a combustion zone where carbon reacts with oxygen to form CO and CO_2. The oxidation provides the overall heat for the gasification and devolatilization reactions which are endothermic. Only a negligible amount of unburned carbon

remains in the ash. Oxidant and steam enter the gasifier near the bottom and are heated as they rise upward to the combustion zone. The oxidant flow is controlled to effect the gasification of the coal and the steam rate is controlled to maintain a specified gasifier bottom temperature.

The crude gas leaving the gasifier contains appreciable quantities of tars, oils, naphtha, phenols, fatty acids, ammonia, hydrogen sulphide, sulphur compounds and a small amount of coal and ash dust. The raw effluent at 581°C flows through the quench scrubber where it is washed with process condensate. The washing process quenches the gas to 193°C and condenses the high boiling tar fractions. Coal and ash dust are removed with the condensed tar leaving the quenched effluent gas essentially free of particulate matter. Ash is continuously removed from the gasifier through the ash lockhopper and discharged in a batchwise manner.

7.1.1.3 Gas Cooling

The raw gasifier effluent exiting the quench scrubber is cooled to 41°C in a series of exchangers in the gas cooling train. Low level heat recovered in the train is used for fuel gas preheating and producing low pressure, saturated steam. In addition, several preheat exchangers for boiler feedwater and deaerator feedwater have been included. Final heat removal is effected in a cooling water exchanger. The cooled gas is then sent to an Ammonia Scrubber where ammonia is removed. Process condensate from the quench scrubber and the gas cooling train is sent to the water treatment section where hydrocarbon liquids and ammonia formed in the gasifier are recovered.

7.1.1.4 Sulphur Removal

The gas leaving the gas cooling train is directed to the Selexol sulphur removal plant for low temperature purification. The processing scheme is the same as used in the Texaco case and results in removal of 93.3% of the sulphur

in the incoming fuel gas. Purified fuel gas exits the top of the Absorber and the sulphur-rich tail gas is sent to the Sulphur Recovery Plant.

7.1.1.5 Sulphur Recovery

Similar to the Texaco case, the Sulphur Recovery Plant is comprised of a Claus Unit and a Beavon Unit for the tail gas treating. Approximately 99.97% of the sulphur fed to the plant is recovered in elemental form. Sulphur recovery for the Lurgi LT purification case is estimated to be 3329 kg/hr or approximately 90.0 percent of the total sulphur entering in the feed coal. The process description is the same as given for the Texaco case.

7.1.1.6 Fuel Gas Saturation and Preheat

The purified fuel gas exiting the top of the Selexol Absorber is directed to a saturator tower where it is heated with hot water. This step has been incorporated, as in the Texaco case, to inhibit NO_x formation in the gas turbine combustor by reducing the firing temperature. The saturated fuel gas is then preheated in the gas cooling train to prevent water condensation prior to fuel gas introduction into the GT combustor.

7.1.1.7 Combined Cycle System

The process flow schematic for the Lurgi/Low Temperature purification case presents the Combined Cycle Plant in simplified form. The basis for the Combined Cycle Plant electrical production has been reviewed in Section 4.

The preheated fuel gas enters the combustor at 160°C and a gas turbine firing temperature of 1200°C was assumed. The exhaust gas exits the gas turbine expander at 587°C and is directed to the heat recovery steam generator. Generation of high and medium pressure steam in the HRSG for the Lurgi low temperature purification case is augmented by supplementary firing of the

hydrocarbon liquids recovered from the gasification product gas. The process of supplemental fuel firing is utilized to increase the amounts of high and medium pressure steam produced by the plant. The HRSG performs the same functions of preheating, steam generation and superheating in a manner similar to the Westinghouse and Texaco cases. The flue gas exits the heat recovery steam generator in the Lurgi case at 171°C.

7.1.1.8 Steam, Boiler Feedwater and Condensate

For the oxygen-blown, Lurgi low temperature purification case, a schematic representation Figure 7.1-2 of the steam, boiler feedwater and condensate production rates for the plant is included.

The process plant steam system is integrated with the combined cycle plant and operates at the following levels:

High Pressure	-	10,099 kPa, 510°C
Medium Pressure	-	2,618 kPa, 510°C
Low Pressure	-	791 kPa, 170°C
Low Pressure	-	446 kPa, 148°C

The Lurgi plant's steam system operates in a manner similar to that for Westinghouse and Texaco but does include differences attributable to the type of gasification system employed. High pressure steam is generated only in the HRSG for the Lurgi case and superheated medium pressure steam is desuperheated and fed to the Lurgi gasifier along with the MP steam produced in the gasifier jacket. At the low pressure level (446 kPa), portion of the steam is superheated in the HRSG and sent to the condensing turbine to produce additional power. In addition, low pressure steam at both levels is used in the process condensate treatment system. The other process systems generate and use steam similarly to Texaco.

TABLE 7.1-2

Case ML

LURGI/LOW TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

ENERGY IN		Total Heat 10^6 KJ/HR
Coal @ 16°C - 100,000 Kg/hr HHV = 28,459 KJ/Kg		2845.9
Gas Turbine combustion Air at 16°C		27.2
Air to Air Separation Plant at 16°C		3.5
TOTAL ENERGY IN		2876.6
Energy Out		
Power		
Gas Turbine Gen Output (Net of Air Comp, Mechanical Losses and Users, Generator Losses)	732.8	
Steam Turbine Gen Output (Net of Mechanical, Generator Losses and Users, ASU Comp and BFW Pumping)	199.1	
Plant Power Consumption	(22.0)	
NET POWER		909.9
Sulphur Product		31.3
Power Block Losses		
(For Gas Turbine Expander, Generator & ST Gen)		52.6
Gasification		
Heat Losses		21.6
Ash-Slurry Stream		18.1

TABLE 7.1-2 (Continued)

Case ML

LURGI/LOW TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

		Total Heat 10^6 KJ/HR
Exhaust or Waste Gas Streams		
From Gas Turbine		698.5
From Air Separation Plant		0.6
Sulphur Recovery Tailgas		10.2
From Deaerator Vent		1.2
Liquid Streams Exiting		
Wastewater		60.8
Ammonia		28.0
Rejection to Cooling Water		
Gas Cooling and Quench	65.3	
Air Sepn. Plant Compression Condensing	99.5	
Selexol Cooling	19.7	
Turbine Condensers	593.9	
Sulphur Recovery	2.6	
Tar/Oil Separation Unit	138.4	
Phenol Recovery Unit	9.3	
Ammonia Recovery	88.3	
NET HEAT REJECTION		1,017.0
Miscellaneous or Unaccountable Losses		26.8
TOTAL ENERGY OUT		2876.6

7.1.2 Lurgi Dry Bottom Gasifier with High Temperature Purification

(Case MH)

Medium Btu gas production based on the oxygen blown Lurgi Dry-Ash Gasification Process is presented on the process flow schematic Figure 7.1-3. Sulphur removal is achieved through high temperature absorption using zinc ferrite and the facility is integrated with a Combined Cycle Plant for the production of electric power.

The facility consists of the following sections:

- Air Separation
- Lurgi Dry-Ash Gasification Process
- Particulate Removal
- High Temperature Sulphur Removal
- Trace Contaminant Removal
- Absorption Bed Regeneration and Sulphuric Acid
- Combined Cycle

Offsite and environmental treatment units required for the facility's operation are also included in the design.

Reference should be made to Table 7.1-3 for the Material Balance of this case and to Table 7.1-4 for the Overall Energy Balance.

7.1.2.1 Air Separation Plant

The description for this section is the same as that given for the Lurgi Gasification/Low Temperature Purification case.

7.1.2.2 Lurgi Dry-Ash Gasification Process

The description for this section is the same as that given for the Low Temperature Purification case except there is not gas scrubbing' the gas passing directly to the particulate and sulphur removal systems at about 580°C.

7.1.2.3 Particulate Removal

Most of the dust in the gas is removed by the cyclone system. The remaining particulate matter is removed by ceramic fiber filters. These systems are described in sections 6.1.2.4 and 6.1.2.3.

7.1.2.4 High Temperature Sulphur Removal
Trace Contaminant Removal
Absorption Bed Regeneration and Sulphuric Acid

The descriptions for these systems are the same as those given for the Texaco Gasification/High Temperature Purification case, except the gas enters the absorption bed at about 580°C and leaves at about 613°C. The sulphur content of the fully loaded zinc ferrite will be less than 15% in this case due to the lower absorption temperature and the large amount of steam present. A figure of 10% has been assumed for this study.

7.1.2.5 Combined Cycle

The basis of the combined cycle design has been given in Section 4. A brief description of the plant follows.

The fuel gas is burned in the gas turbine combustor and the combustion products enter the gas turbine expander at 1200°C. The high steam content in the gas will provide good NO_x control. The exhaust gas leaves the gas

turbine at about 592°C and passes to the heat recovery system where it performs the following functions:

- Preheating of boiler feedwater deaerator feedwater
- Generation of high, medium and low pressure steam
- Superheating of high pressure steam and reheating of medium pressure steam.

The flue gas leaves the heat recovery system at about 108°C.

7.1.2.6 Steam, Boiler Feedwater and Condensate

The steam, feedwater and condensate flow rates and conditions are shown on the steam system schematic, Figure 7.1-4.

The system is similar to that described for the Texaco Gasification/High Temperature Purification case except:

- There is no gasification boiler
- The gasifier consumes medium pressure steam
- All medium pressure steam is superheated to 510°C
- Medium pressure steam is at 2618 kPa.

TABLE 7.1-4

Case MH

LURGI/HIGH TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

ENERGY IN		Total Heat 10^6 KJ/HR
Coal @ 16°C - 100,000 Kg/hr HHV = 28,459 KJ/Kg		2845.9
Gas Turbine Combustion Air at 16°C		36.3
Air to Air Separation Plant at 16°C		3.5
TOTAL ENERGY IN		2885.7
ENERGY OUT		
Power		
Gas Turbine Generator	987.5	
Steam Turbine General Output (Net of ASU Compression and BFW Pumping Steam Requirements)	182.2	
Plant Power Consumption	(37.4)	
NET POWER		1132.3
Sulphuric Acid Product		(58.1)
Power Block Losses		
(For Gas Turbine Expander, Generator and ST Gen)		64.8
Gasification		
Heat Losses		23.7
Ash Stream		18.1
Exhaust or Waste Gas Stream		
From Gas Turbine		952.8
From Air Separation Plant		0.6
From Deaerator Vent		1.2

TABLE 7.1-4 (Continued)

Case MH

LURGI/HIGH TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

		Total Heat 10^6 KJ/HR
Rejection to Cooling Water or Air		
Air Separation Plant Compression Cooling	99.5	
Sulphuric Acid Plant	53.3	
Turbine Condensers	536.0	
Sulphur Absorption Bed Regeneration	75.5	
NET HEAT REJECTION		764.3
Miscellaneous or Unaccountable Losses		14.0
TOTAL ENERGY OUT		2885.7

7.1.3 Westinghouse Fluidized Bed Gasifier with Low Temperature Purification

(Case FL)

Medium BTU gas production based on the oxygen-blown Westinghouse gasification process is presented on the attached process flow schematic. Sulphur removal is effected through the low temperature Selexol Process and the facility is integrated with a combined cycle plant. Similar to the Texaco case, 100 metric tonnes per hour of Illinois No. 6 coal is gasified into MBG fuel ultimately to be used for power generation.

The facility consists of the following sections:

- o Air Separation
- o Coal Preparation and Drying
- o Westinghouse Gasification
- o Gas Cooling
- o Selexol Sulphur Removal
- o Sulphur Recovery Plant
- o Combined Cycle Plant

Additionally, necessary offsite and environmental treatment units required for the plant's operation are included.

Reference can be made to Table 7.1-5 for the Material Balance of this csase and to Table 7.1-6 for the Overall Energy Balance.

7.1.3.1 Coal Preparation and Drying

As received coal is ground to minus 6.35 mm x 0 mm size and, if necessary, dried to a free flowing condition. The coal is transferred in a batch-wise operation from the atmospheric pressure, coal surge bin to the pressurized coal lockhoppers. The coal feed lockhoppers operate at a constant pressure which is sufficient for pneumatic conveying of the coal to the gasifier.

Coal is fed continuously from the coal feed lockhoppers to the gasifiers through a pneumatic transfer line in which recycled product gas is used as the transport medium.

7.1.3.2 Air Separation Plant

The oxidant feed system provides 98.5 mole percent oxygen to the gasification system at an absolute pressure of approximately 3300 kPa. The plant, whose operation has been described in the Texaco case, provides almost 62 metric tonnes per hour to the Westinghouse gasifiers.

7.1.3.3 Westinghouse Coal Gasification Process

The Westinghouse Coal Gasification Process is a proprietary system which includes a reaction vessel, a heat and particulate recovery system and a gas quench system. The gasifier is a single stage, pressurized fluidized bed reactor. The vessel is cylindrical, vertically oriented, refractory lined and constructed of carbon steel. Coal, steam and oxygen react within the gasifier to produce a medium Btu gas which exits near the top of the vessel. Ash waste is discharged from the bottom.

The coal, steam and oxidant enter through a central feed system located in the ash separator. Recycled char fines and transport gas coupled with the other feed streams emerge from the feed system to form a jet where the coal and oxidant react. The coal is quickly devolatilized and partially combusted. The resulting solid product, char, forms the gasifier bed material which is fluidized by the steam, oxidant, and recycle gas streams. The reactor design is based upon proprietary relationships which consider coal properties, fluidization requirements and reaction kinetics for the gasification and coal combustion processes.

section before exiting the reactor. The freeboard area reduces particle entrainment in the raw product gas and allows additional time for reactions.

The Westinghouse plant steam system's operation is similar to the Texaco operation with some minor differences. Medium pressure steam is desuperheated and used in the gasification process, and some of the low pressure steam is used for fuel gas superheating and for process condensate treatment. The other process systems generate and use steam in a manner similar to Texaco.

Ash withdrawal is achieved through a series of lockhoppers located beneath each gasifier. The system is designed to receive agglomerated dry ash particles, depressurize and store them temporarily prior to ultimate disposal. The lockhoppers operate at gasification and ambient pressure to receive and discharge ash in a batchwise manner.

The heat/particulate recovery system, located downstream of the gasifier consists of a cyclone followed by a series of heat exchangers. Entrained particles are removed from the hot product gas by the cyclone and are then pneumatically conveyed to the gasifier using recycled product gas as the transport medium. Sensible heat is recovered from the product gas in high, medium, and low pressure waste heat boilers which increases overall process efficiency. The gas exiting the low pressure boiler is further cooled and scrubbed of any remaining particulate matter entrained in the gas by a direct water quench venturi scrubbing system. The system produces a saturated, essentially particulate free raw gas which is sent to the gas cooling train, and a waste slurry stream for disposal. The saturated raw gas is at 121°C.

7.1.3.4 Gas Cooling Train

Low level heat is recovered in the gas cooling train with the heat utilized for various preheat services. Subsequent to water cooling to approximately 38°C, the raw gas enters the Sour Water Scrubber where ammonia and a small amount of sulphur compounds are removed. The sour process liquid is sent to

the Condensate Treatment Unit and the product gas is split into two streams; one to downstream sulphur processing and the other used as recycled gas.

7.1.3.5 Sulphur Removal

The Selexol process is used for selective sulphur removal in the oxygen-blown, Westinghouse low temperature purification case. The processing scheme is the same as used in the Texaco case for low temperature purification and results in removal of 89.2% of the sulphur from the incoming fuel stream. The purified gas exits the top fo the Absorber and the sulphur rich tail gas from the regenerator is sent to the Sulphur Recovery Plant.

7.1.3.6 Sulphur Recovery

The Sulphur Recovery is comprised of a Claus Unit and a Beavon Unit for tail gas treating. The sulphur-rich tail gas from the sulphur removal plant and an off gas from the condensate treatment system are combined and fed to Sulphur Recovery. Approximately 99.97% of the sulphur fed to the plant is recovered in elemental form.

Sulphur recovery for the Westinghouse LT purification case is estimated to be 3332 kg/hr or approximately 90.1 percent of total sulphur entering in the feed coal. The process description is the same as given for the Texaco case.

7.1.3.7 Fuel Gas Saturation and Preheat

The purified fuel gas exiting the top of the Selexol Absorber is directed to a saturator tower where it is heated with hot water. This step has been incorporated, as in the Texaco case, to inhibit NO_x formation in the gas turbine combustor by reducing the firing temperature. The saturated fuel gas is then preheated with low pressure steam to prevent water condensation prior to fuel gas introduction into the GT combustor.

7.1.3.8 Combined Cycle System

The process flow schematic for the Westinghouse/Low Temperature purification case is similar to the Texaco case for power generation and heat recovery. The basis has been described in Section 4.

The preheated fuel gas enters the combustor at 163°C and a gas turbine firing temperature of 1200°C was assumed. The exhaust gas exits the turbine expander at 584°C and is sent to the heat recovery steam generator where boiler feedwater preheating, high and medium pressure steam generation and superheating are conducted. The flue gas exits the heat recovery steam generator at 166°C.

7.1.3.9 Steam, Boiler Feedwater and Condensate

The steam, boiler feedwater and condensate production rates for the plant for the Westinghouse low temperature purification case are presented on the schematic diagram, Figure 7.1-6.

The process plant steam system is integrated with the combined cycle plant and operates at the following levels:

High Pressure	-	10,099 kPa, 510°C
Medium Pressure	-	3,032 kPa, 357°C
Low Pressure	-	791 kPa, 170°C
Low Pressure	-	446 kPa, 148°C

The Westinghouse plant stream system's operation is similar to the Texaco operation with some minor differences. Medium pressure steam is desuperheated and used in the gasification process, and some of the low pressure steam is used for fuel gas superheating and for process condensate treatment. The other process systems generate and use steam in a manner similar to Texaco.

TABLE 7.1-6

Case FL

WESTINGHOUSE/LOW TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

ENERGY IN		Total Heat 10^6 KJ/HR
Coal @ 16°C - 100,000 Kg/hr HHV = 28,459 KJ/Kg		2845.9
Gas Turbine combustion Air at 16°C		29.9
Air to Air Separation Plant at 16°C		4.4
TOTAL ENERGY IN		2880.2
Energy Out		
Power		
Gas Turbine Gen Output (Net of Air Comp, Mechanical Losses and Users, Generator Losses)	732.8	
Steam Turbine Gen Output (Net of Mechanical, Generator Losses and Users, ASU Comp and BFW Pumping)	282.1	
Plant Power Consumption	(36.0)	
NET POWER		1020.0
Sulphur Product		31.3
Power Block Losses		
(For Gas Turbine Expander, Generator & ST Gen)		58.0
Gasification		
Heat Losses		21.6
Ash-Slurry Stream		15.8

TABLE 7.1-6 (Continued)

Case FL

WESTINGHOUSE/LOW TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

		Total Heat 10^6 KJ/HR
Exhaust or Waste Gas Streams		
From Gas Turbine		664.0
From Air Separation Plant		1.3
Sulphur Recovery Tailgas		5.6
From Deaerator Vent		1.2
Liquid Streams Exiting		
Wastewater		8.7
Ammonia		4.1
Rejection to Cooling Water or Air		
Gas Cooling Train	21.5	
Air Sep Plant Compression Condensing	119.7	
Selexol Cooling	10.4	
Turbine Condensers	708.7	
Sulphur Recovery	2.6	
Process Condensate Treatment	11.6	
Quench System	125.0	
NET HEAT REJECTION		999.5
Miscellaneous or Unaccountable Losses		69.1
TOTAL ENERGY OUT		2880.2

7.1.4 Westinghouse Fluidized Bed Gasifier with High Temperature Purification

(Case FH)

Medium BTU gas production based on the oxygen-blown Westinghouse gasification process is presented on the attached process flow schematic Figure 7.1-7. Sulphur removal is effected through the high temperature absorption using zinc ferrite and the facility is integrated with a combined cycle plant for power generation.

The facility consists of the following sections:

- o Air Separation
- o Coal Preparation and Drying
- o Westinghouse Gasification and Heat Recovery
- o Particulate Removal
- o Recycle Gas Cooling
- o High Temperature Sulphur Removal
- o Trace Contaminant Removal
- o Absorption Bed Regeneration and Sulphuric Acid
- o Combined Cycle Plant

Additionally, offsite and environmental treatment systems required for the plant's operation are included in the design.

Reference can be made to Table 7.1-7 for the Material Balance of this case and to Table 7.1-8 for the Overall Energy Balance.

7.1.4.1 Air Separation Plant

7.1.4.2 Coal Preparation and Drying

The descriptions for this section is the same as that given for the Westinghouse Gasification Low Temperature Purification case, except the gas is

cooled to about 650°C in the high pressure boiler before passing to the ceramic fiber filters. There is no scrubbing stage.

7.1.4.3 Particulate Removal

Most of the ash and char is removed from the raw gas by the cyclone, which is an integral part of the gasification package. The remaining particulate matter is removed by ceramic fibre filters. These are described in Section 6.1.2.3.

7.1.4.4 Recycle Gas Cooling

The recycle gas required by the gasification process is cooled in a series of heat exchangers before compression. These generate high pressure steam and heat high pressure boiler feedwater and deaerator feedwater. The gas is finally cooled with cooling water and process condensate removed before passing to the recycle gas compressor.

7.1.4.5 High Temperature Sulphur Removal

7.1.4.6 Trace Contaminant Removal

7.1.4.7 Absorption Bed Regeneration and Sulphuric Acid

The descriptions for these systems are the same as those given for the Texaco Gasification/High Temperature Purification case, except the gas leaves the absorption bed at an average temperature of about 706°C.

7.1.4.8 Combined Cycle

The basis of the combined cycle design has been detailed in Section 4. A brief description of the plant is as follows.

The fuel gas is burned in the gas turbine combustor, and the combustion products enter the gas turbine expander at 1200°C. It is assumed that the steam content in the fuel will provide adequate NO_x control. The exhaust gas exits the gas turbine expander at 582°C and is sent to the heat recovery steam generator where the following fuctions are performed:

- Preheating of high and medium pressure boiler feedwater and deaerator feedwater.
- Generation of high, medium and low pressure steam.
- Superheating of high pressure and medium pressure steam.

The flue gas leaves the heat recovery system at about 104°C.

7.1.4.9 Steam, Boiler Feedwater, and Condensate

The steam, feedwater and condensate flow rates and conditions are shown on the attached steam system schematic, Figure 7.1-8.

The system is similar in principle to that described for the Texaco Gasification/High Temperature Purification case except medium pressure steam is consumed by the gasification process and heat from the recycle gas cooling train is used to generate HP steam and preheat feedwater.

TABLE 7.1-8

Case FH

WESTINGHOUSE/HIGH TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

		Total Heat 10^6 KJ/HR
ENERGY IN		
Coal @ 16°C - 100,000 Kg/hr HHV = 28,459 KJ/Kg		2845.9
Gas Turbine Combustion Air at 16°C		32.0
Air to Air Separation Plant at 16°C		4.4
TOTAL ENERGY IN		2882.3
ENERGY OUT		
Power		
Gas Turbine Generator	836.2	
Steam Turbine General Output (Net of ASU Compression and BFW Pumping Steam Requirements)	294.3	
Plant Power Consumption	(51.5)	
NET POWER		1079.0
Sulphuric Acid		(58.1)
Power Block Losses		
(For Gas Turbine Expander, Generator and ST Gen)		63.2
Gasification		
Heat Losses		14.6
Ash-Slurry Stream		15.8
Exhaust or Waste Gas Streams		
From Gas Turbine		513.5
From Air Separation Plant		1.3
From Deaerator Vent		1.2

TABLE 7.1-8 (Continued)

Case FH

WESTINGHOUSE/HIGH TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

		Total Heat 10^6 KJ/HR
Wastewater		1.0
Rejection to Cooling Water		
Gas Cooling Train	1.8	
Air Sepn. Plant Compression Condensing	119.7	
Sulphuric Acid Plant	53.3	
Turbine Condensers	852.1	
Sulphur Absorption Bed Regeneration	75.5	
Process Condensate Treatment	6.8	
Quench System	13.9	
NET HEAT REJECTION		1123.1
Miscellaneous or Unaccountable Losses		127,7
TOTAL ENERGY OUTPUT		2882.3

7.1.5 Texaco Entrained Bed Gasifier with Low Temperature Purification

(Case EL)

A gas roots facility for medium Btu gas production based on the oxygen-blown, Texaco gasification process employing low temperature sulphur removal and integrated with a combined cycle plant is shown on the process flow schematic Figure 7.1-9. All calculations were based on a fixed coal consumption of 100 metric tonnes/hr of Illinois No. 6 coal to arrive at a net electric power for export from a combined cycle facility. The main sections of the processing facility consists of an air separation unit for oxygen production, gasification, gas cooling, sulphur removal units and a combined cycle power system. In addition to the main processing units, the facility includes a slurry preparation section, a sulphur recovery and tailgas treatment system and other necessary, off-site and environmental units required for the plant's operation.

Reference can be made to Table 7.1-9 for the Material Balance of this case and to Table 7.1-10 for the Overall Energy Balance.

7.1.5.1 Slurry Preparation

Sized coal (38 MM x 0) is withdrawn from storage silos by vibrating feeders and transported by a belt conveyor or the slurry preparation section. After mixing with recycle solids and water, the coal is ground in a wet milling process and transferred to agitated, coal slurry run tanks. Agitation is necessary to keep the solids suspended and to provide a uniform coal slurry to the gasifier. A transfer pump is used to withdraw the slurry from the run tank and feed it to the high pressure charge pumps for charging the gasifiers. The water slurry fed to the gasifiers contains approximately 66.5 weight percent coal.

7.1.5.2 Air Separation Plant

The oxidant feed system provided 98 mole percent oxygen to the gasification process at 149°C and and absolute pressure of 5066 kPa. In this system, atmospheric air is compressed to 763 kPa in a two state centrifugal machine and subsequently processed in an air separation unit. This unit cryogenically separates oxygen from the constituents in air and produces oxygen (of 98 percent purity) at 115 kPa pressure and 32°C. The oxygen product is compressed in a multistage centrifugal compressor and directed to the Texaco gasifier.

7.1.5.3 Texaco Coal Gasification Process

The Texaco Coal Gasification system includes the coal gasifier, fuel gas cooling and finally particulate removal in the carbon scrubbing unit. The gasifier is a vertical, cylindrical, refractory lined vessel with a carbon steel shell. Coal slurry and oxygen combine at the gasifier burners and move downward from the top of the vessel. Partial combustion with oxygen takes place at a pressure of 4238 kPa and a temperature in the range of 1260°C to 1427°C to produce gas consisting mainly of CO, H_2, CO_2 and steam. Most of the sulphur in the coal is converted to H_2S and the balance is converted to COS. Nitrogen and argon from the oxygen feed, along with most of the nitrogen from the coal, appear in the gas. The gas contains small amounts of methane, ammonia, some unconverted carbon and the ash not removed in the form of slag.

The ash in the coal feed melts to form slag in the gasifier. The temperature in this section must be above the ash flow point to ensure free flowing molten slag. The bulk of molten slag will solidify in the radiant cooler and drop into a lockhopper at the bottom of the cooler. The slag is ultimately removed from the system through the lockhopper system.

The raw, hot fuel gas issuing from the gasifier is cooled by the radiant and convection coolers generating high pressure (10,480 kPa) saturated steam in the process. The temperature of the gas is well below the ash softening point thereby solidifying the entrained ash particules. The gas then enters the carbon scrubber where most of the fine particulate material is removed. Counter current contact with water captures entrained particles and produces an essentially particulate-free, saturated gas. The temperature at this point is approximately 178°C.

7.1.5.4 Gas Cooling Train

The gas exiting the Carbon Scrubber is further cooled in a gas cooling train to 38°C. The low level heat recovered in this section is used for various preheat and reheat services integrated with otherplant sections. The water condensate produced by the cooling process is directed to the water treatment system for purification and treatment.

7.1.5.5 Sulphur Removal

The sulphur removal system employs the Selexol Process for the selective removal of sulphur compounds (primarily H_2S) from the fuel gas stream. The sulphur compounds are absorbed in the Selexol solvent in order to remove 90 percent of sulphur from the incoming fuel stream.

The raw gas from the gas cooling train is mixed with a recycle flash gas and fed to the bottom of the Absorber, where it contacts Selexol solvent counter-currently in a packed bed. The purified gas exits the top of the Absorber at 41°C. The rich solvent from the bottom of the Absorber is flashed to low pressure with the flash gas recycled to the raw gas feed stream and the flash liquid fed to the Regenerator.

In the Regenerator, the absorbed sulphur compounds (H_2S and COS) and carbon dioxide are stripped from the solvent. Reboil heat is provided by steam in a

vertical thermosyphon reboiler. The hot, lean solution from the bottom of the Regenerator is cooled by interchange with the Regenerator feed stream, cooling water and boiler feed water before entering the top of the Absorber.

The overhead from the top of the Regenerator is cooled by preheating boiler feed water, condensing water which provides reflux to the tower. The sulphur rich tailgas produced is sent to the Sulphur Recovery Plant for final processing.

7.1.5.6 Sulphur Recovery

The Sulphur Recovery Plant for the facility consists of a Claus Sulphur Recovery Unit and a Beavon Unit for tailgas treating. By use of these process systems approximately 99.96% of the sulphur in the sulphur-rich tailgas from the Selexol plant is recovered in elemental form. Sulphur recovery for the Texaco, low temperature purification is estimated to be 3329 kg/hr or approximately 90 percent of total sulphur entering in the feed coal.

In the Claus unit, the Selexol tailgas and a stoichiometric quantity of air are fed into a reaction furnace where one third of the H_2S is burned to SO_2. The H_2S and SO_2 react to form elemental sulphur thermally in the furnace and catalytically in a series of reactors according to the Claus reaction:

$$2H_2S + SO_2 = 3S + 2H_2O + \text{Heat}$$

The reaction which is exothermic is catalyzed by a bauxite or alumina catalyst in packed bed reactors. The sulphur produced is recovered by cooling the effluent below the sulphur dew point with boiler feed water in sulphur condensers. The heat of reaction is recovered in the sulphur condensers through the production of medium pressure steam. The condensed sulphur flows by gravity to a concrete sulphur sump which is kept in a molten state by condensing low pressure steam.

The Claus unit produces a tailgas from the last reactor containing unreacted H_2S and CO_2. To meet environmental emissions limits, the tailgas is processed further in a Beavon tailgas treating unit to remove these sulphur compounds.

The Beavon unit employs a proprietary design which is a modification of the Stretford process. In the Beavon-Stretford process, sulphur compounds present in the Claus effluent gas are converted to H_2S in a catalytic reactor according to the following hydrogenation and hydrolysis reactions:

$$SO_2 + 3H_2 \longrightarrow H_2S + 2H_2O$$
$$S_8 + 8H_2 \longrightarrow 8H_2S$$
$$COS + H_2O \longrightarrow CO_2 + H_2S$$
$$CS_2 + 2H_2O \longrightarrow CO_2 + 2H_2S$$
$$CO + H_2O \longrightarrow CO_2 + H_2$$

These reactions result in the conversion of essentially all sulphur compounds to hydrogen sulphide and a small residual amount of carbonyl sulphide. Removal of H_2S from the gas stream is accomplished in the Stretford plant absorber by direct contact with a lean Stretford solution at atmospheric pressure. The effluent gas exiting the absorber is vented directly to the atmosphere and the sulphur is recovered in elemental form. The recovered sulphur is deposited in the Claus Plant sulphur sumps and the Stretford solution is regenerated with air for further use.

7.1.5.7 Gas Expansion, Saturation and Preheat

The saturated fuel gas exiting the top of the Selexol absorber is preheated in the gas cooling train and subsequently expanded in the Fuel Gas Expander to recover additional power prior to combustion. The expanded gas is sent to a saturator tower where it is heated with hot water. The saturated fuel gas is preheated again in the cooling train to prevent water condensation prior to fuel gas introduction into the combustor of the gas turbine.

The saturator tower is included in the plant design for the express purpose of inhibiting NO_x formation in the gas turbine combustor. The NO_x formed by combustion is a function of complex factors and maybe limited through design features, incorporated into the combustor and steam or water introduction into the fuel gas to reduce the firing temperature. Through the use of a heater-saturator loop, the fuel gas is heated and saturated at a predetermined temperature to ensure that the requisite amount of water vapor for NO_x limitation is added. This design has been included not to optimally solve the NO_x formation problem but to address it with a viable solution.

7.1.5.8 Combined Cycle System

The process flow schematic for the Texaco/Low Temperature sulphur removal case depicts, in simple form, a combined cycle plant for power generation and heat recovery. The basis for the Combined Cycle Plant electrical production has been reviewed in detail in section 4. For completeness, a brief process review of the plant follows.

The fuel gas, preheated to prevent water condensation, enters the combustor at 160°C. A gas turbine firing temperature of 1200°C was used for calculations involving combined cycle plant performance. The exhaust gas exists the gas turbine expander at 585°C and is sent to the heat recovery steam generator where the following functions are performed:

- Preheating of high pressure and medium pressure boiler feed water in separate coils.

- Generation of high pressure and medium pressure steam in separate coils.

- Superheating of high pressure and medium pressure steam in separate coils.

The flue gas exits the heat recovery steam generator at 162°C, well above the temperature at which corrosion problems may exist for such systems.

7.1.5.9 Steam, Boiler Feedwater, and Condensate

Included for the oxygen-blown Texaco, low temperature purification case is a schematic representation Figure 7.1-10 of the steam, boiler feedwater and condensate production rates for the plant.

The process plant steam system is integrated with the combined cycle plant and operates at the following levels:

High Pressure	-	10,099 kPa, 510°C
Medium Pressure	-	3,032 kPa, 357°C
Low Pressure	-	791 kPa, 170°C
Low Pressure	-	446 kPa, 171°C

High pressure steam generation is carried out in the gasification waste heat boilers and in the heat recovery system associated with the combustion turbine. The saturated, nigh pressure steam generated is then superheated to 510°C in the HRSG superheater. All the superheated H.P. steam is used to drive a single back-pressure type turbine. The steam enters the turbine at 510°C and a pressure of 10,099 kPa and exhausts at 3032 kPa.

Medium pressure steam is generated in the gasification and sulphur recovery unit waste heat boilers, in the heat recovery steam generator and by cooling gas turbine combustor bypass air. This medium pressure steam is superheated to 427°C in the HRSG's steam reheater and mixed with exhaust steam from the high pressure steam turbine. The superheated M.P. steam at 357°C and 3032 kPa is then used in a condensing steam turbine. A small portion of the steam flow in the turbine is extracted at 894 kPa (238°C), desuperheated and used for various process services at 991 kPa and 170°C. A secondary low pressure steam level at 446 kPa and 171°C is generated by waste heat boilers in the gasification, gas cooling, and sulphur recovery units. A small portion of the steam generated is used in heating the sulphur pit (keeps sulphur in a molten state) with the balance used for deaerating boiler feedwater.

Raw boiler feedwater is treated in a demineralization unit and held in a storage tank until make-up BFW (Boiler Feed Water) required. The make-up is preheated in the lean Selexol cooler and the Selexol overhead condenser of the Sulphur Removal Plant and fed to the deaerator. Condensate from the steam turbine condenser is preheated in the gas cooling train condensate heater to 106°C and routed to the deaerator. The deaerator operates at 135°C. Deaerated boiler feedwater is pumped to high pressure, preheated in the low temperature HRSG economizer, and ultimately provides the feedwater required for medium and high pressure steam level production. Subsequent to low level preheat in the first economizer, high pressure BFW is preheated to 288°C in the high temperature HRSG economizer and then sent to the high pressure boilers.

TABLE 7.1-10

Case EL

TEXACO/LOW TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

ENERGY IN		Total Heat 10^6 KJ/HR
Coal @ 16°C - 100,000 Kg/hr HHV = 28,459 KJ/Kg		2845.9
Gas Turbine combustion Air at 16°C		29.0
Air to Air Separation Plant at 16°C		6.0
TOTAL ENERGY IN		2881.5
Energy Out		
Power		
Gas Turbine Gen Output (Net of Air Comp, Mechanical Losses and Users, Generator Losses)	754.7	
Steam Turbine Gen Output (Net of Mechanical, Generator Losses and Users, ASU Comp and BFW Pumping)	316.2	
Plant Power Consumption	(34.2)	
NET POWER		1036.7
Sulphur Product		31.3
Power Block Losses		
(For Gas Turbine Expander, Generator & ST Gen)		57.3
Gasification		
Heat Losses		14.5
Ash-Slurry Stream		22.6

TABLE 7.1-10 (Continued)

Case EL

TEXACO/LOW TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

		Total Heat 10^6 KJ/HR
Exhaust or Waste Gas Streams		
From Gas Turbine		592.3
From Air Separation Plant		1.8
Sulphur Recovery Tailgas		3.3
From Deaerator Vent		1.2
Rejection to Cooling Water		
Gas Cooling Train	7.0	
Air Sep Plant Compression Condensing	168.8	
Selexol Cooling	14.8	
Turbine Condensers	912.8	
Sulphur Recovery	2.6	1106.0
NET HEAT REJECTION		
Miscellaneous or Unaccountable Losses		13.9
TOTAL ENERGY OUT		2880.9

7.1.6 Texaco Entrained Bed Gasifier with High Temperature Purification

(Case EH)

Medium BTU gas production based on the oxygen-blown Texaco gasification process is presented on the process flow schematic Figure 7.1-11. Sulphur removal is effected through high temperature absorption using zinc ferrite, and the facility is integrated with a combined cycle plant for power generation.

The facility consists of the following sections:

- Air Separation
- Slurry Preparation
- Texaco Gasification and Heat Recovery
- Particulate Removal
- High Temperature Sulphur Removal
- Trace Contaminant Removal
- Absorption Bed Regeneration and Sulphuric Acid
- Combined Cycle

Additionally, offsite and environmental systems required for plant operation are included in the design.

Reference can be made to Table 7.1-11 for the Material Balance of this case and to Table 7.1-12 for the Overall Energy Balance.

7.1.6.1 Slurry Preparation

7.1.6.2 Air Separation

The descriptions for these sections are the same as those given for the Texaco Gasification/Low Temperature Purification case.

7.1.6.3 Texaco Coal Gasification and Heat Recovery

The description for this section is the same as that given for the Low Temperature Purification case, except the gas is cooled to about 650°C in the radiant cooler before passing to the particulate removal stage. There is no convection cooler or carbon scrubber.

7.1.6.4 Particulate Removal

Most of the slag and char leaving the gasifier is removed from the system through the lockhopper at the base of the radiant cooler. The remaining particulate matter is removed in a combination of cyclones and ceramic fiber filters. These systems are described in detail in Sections 6.1.2.4 and 6.1.2.3.

7.1.6.5 High Temperature Sulphur Removal

Following particulate removal, the gas enters the zinc ferrite sulphur absorption bed at 650°C. The zinc ferrite absorbs sulphur to about 15% of its initial weight. During the absorption process part of the carbon monoxide in the gas is hydrolysed to hydrogen. Small amounts of hydrogen and carbon monoxide are oxidized to carbon dioxide and water during the reduction of the iron oxide. The gas leaves the absorption bed at an average temperature of 727°C. The temperature rise through the bed is largely the result of the carbon monoxide hydrolysis reaction.

Further information on zinc ferrite sulphur absorption is given in Section 6.2.2.3.

7.1.6.6 Trace Contaminant Removal

For the purposes of this study it is assumed that a technology such as that being developed by Westinghouse utilizing an alumino silicate getter will

provide adequate protection for the combustion turbines. This technology is still under development.

7.1.6.7 Absorption Bed Regeneration System

After an absorption period of about 4 hours, the sulphur containing raw gas is passed to a bed of freshly regenerated sorbent, and the sulphur-laden bed is connected to the regeneration system. Compressed air and recycle gas (containing primarily nitrogen and sulphur dioxide) are preheated in the feed/effluent exchanger before passing into the zinc ferrite bed at about 800°C. The oxygen in the fed is rapidly consumed, zinc and iron sulphides being converted to their oxides with the release of sulphur dioxide. The hot effluent gases are used to preheat the feed. Part of the effluent passes to the sulphuric acid plant where at least 98% of the sulphur dioxide is recovered as 98% sulphuric acid. The remainder is cooled to about 40°C before recompression and recycle. The regeneration phase takes about 8 hours, so for each bed in absorption mode there will be 2 beds at different stages of regeneration. This will reduce the fluctuation of sulphur dioxide flow rate to the acid plant.

A recycle gas/air ratio of about 9.1 is required to keep the bed temperature below 1000°C during regeneration. It would probably be necessary to fluidize the bed during regeneration to prevent excessive temperature excursions within the bed. The mechanical implications of this have not been addressed though allowance has been made for the power requirements for such a system.

7.1.6.8 Fuel Gas Expansion and Combined Cycle

The clean fuel gas from the trace contaminant removal system passes into a fuel gas expander at about 725°C, 4000 kPa. The expanded gas then passes to the combustion turbine system at about 600°C, 2000 kPa. The basis of the combined cycle design has been detailed in Section 4. A brief description of the plant is as follows.

The fuel gas is burned in the gas turbine combustor, and the combustion products enter the gas turbine expander at 1200°C. It is assumed that the steam content in the fuel will provide adequate NO_x control. The exhaust gas exits the gas turbine expander at 583°C and is sent to the heat recovery steam generator where the following fuctions are performed:

- Preheating of high pressure and medium pressure boiler feedwater and deaerator feedwater.
- Generation of high pressure, and low pressure steam.
- Superheating of high pressure and medium pressure steam.

The flue gas exits the heat recovery steam generator at about 98°C.

7.1.6.9 Steam, Boiler Feedwater, and Condensate

The steam, boiler feedwater and condensate production rates and conditions are shown on the attached schematic.

The process plant steam system is integrated with the combined cycle plant and operates at the following levels:

High Pressure	-	10,099, 510°C
Medium Pressure	-	3,032 kPa, 357°C
Low Pressure	-	466 kPa, 148°C

High pressure steam generation is carried out in the gasification waste heat boiler and in the heat recovery system (HRSG) associated with the combustion turbine. The saturated, high pressure steam generated is then superheated to 510°C in the HRSG superheater. All the superheated H.P. steam is used to drive a back-pressure type turbine. The steam enters the turbine at 510°C and a pressure of 10,099 kPa and exhausts at 3032 kPa, 349°C.

Medium pressure steam is generated in the heat recovery steam generator and by cooling gas turbine combustor bypass air. This medium pressure steam is superheated to 427°C in the HRSG and mixed with exhaust steam from the high pressure steam turbine. The superheated M.P. steam at 357°C and 3032 kPa is then used in a condensing steam turbine. Low pressure steam at 446 kPa and 148°C is generated by waste heat boilers in the HRSG and sulphuric acid plant. Part of the low pressure steam generated is used for deaerating boiler feedwater, the remainder being inducted into the condensing steam turbine.

Raw boiler feedwater is treated in a demineralization unit and held in a storage tank until make-up BFW is required. Condensate from the steam turbine condenser is preheated in the HRSG deaerator feedwater heater to 124°C. The deaerator operates at 135°C. Deaerated boiler feedwater is pumped to high pressure, preheated in the HRSG MP economizer, and ultimately provides the feedwater required for medium and high pressure steam level production. The high pressure BFW is then preheated to 288°C in the HRSG HP economizer and then sent to the high pressure boilers.

7.1.6.10 Particulate Removal

Most of the dust in the gas is removed by the cyclone system. The remaining particulate matter is removed by ceramic fiber filters. These systems are described in Sections 6.1.2.4 and 6.1.2.3.

7.1.6.11 High Temperature Sulphur Removal

7.1.6.12 Trace Contaminant Removal

7.1.6.13 Absorption Bed Regeneration and Sulphuric Acid

The descriptions for these systems are the same as those given for the Texaco Gasification/High Temperature Purification case, except the gas enters the

absorption bed at about 580°C and leaves at about 613°C. The sulphur content of the fully loaded zinc ferrite will be less than 15% in this case due to the lower absorption temperature and the large amount of steam present. A figure of 10% has been assumed for this study.

7.1.6.14 Combined Cycle

The basis of the combined cycle design has been given in Section 4. A brief description of the plant follows.

The fuel gas is burned in the gas turbine combustor and the combustion products enter the gas turbine expander at 1200°C. The high steam content in the gas will provide good NO_x control. The exhaust gas leaves the gas turbine at about 592°C and passes to the heat recovery system where it performs the following functions:

- Preheating of boiler feedwater and deaerator feedwater.
- Generation of high, medium and low pressure steam.
- Superheating of high pressure steam and reheating of medium pressure steam.

The flue gas leaves the heat recovery system at about 108°C.

7.1.6.15 Steam, Boiler Feedwater and Condensate

The steam, feedwater and condensate flow rates and conditions are shown on the attached schematic.

The system is similar to that described for the Texaco Gasification/High Temperature Purification case except:

- There is no gasification boiler.
- The gasifier consumes medium pressure steam.
- All medium pressure steam is superheated to 510°C.
- Medium pressure steam is at 2618 kPa.

TABLE 7.1-12

Case EH

TEXACO/HIGH TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

ENERGY IN		Total Heat 10^6 KJ/HR
Coal @ 16°C - 100,000 Kg/hr HHV = 28,459 KJ/Kg		2845.9
Gas Turbine Combustion Air at 16°C		30.6
Air to Air Separation Plant at 16°C		6.0
TOTAL ENERGY IN		2882.5
ENERGY OUT		
Power		
Gas Turbine Generator Output	844.9	
Steam Turbine Generator (Net of ASU Compression and BFW Pumping Steam Requirements)	318.7	
Plant Power Consumption	(48.7)	
NET POWER		1114.9
Sulphuric Acid		(58.1)
Power Block Losses		
(For Gas Turbine Expander, Generator and ST Gen)		61.7
Gasification		
Heat Losses		7.3
Ash-Slurry Stream		22.6
Exhaust or Waste Gas Streams		
From Gas Turbine		451.9
From Air Separation Plant		1.8
From Deaerator Vent		1.2

TABLE 7.1-12 (Continued)

Case EH

TEXACO/HIGH TEMPERATURE
GAS PURIFICATION
OVERALL ENERGY BALANCE
DATUM 16°C, H_2O (LIQ)

		Total Heat 10^6 KJ/HR
Rejection to Cooling Water or Air		
Air Separation Plant Compression Condensing	168.8	
Sulphuric Acid Plant	53.3	
Turbine Condensers	1004.3	
Sulphur Absorption Bed Regeneration	75.5	
NET HEAT REJECTION		1301.9
Miscellaneous or Unaccountable Losses		(22.7)
TOTAL ENERGY OUT		2882.5

7.2 Estimated Capital Cost

A consistent set of criteria for estimating capital requirements and cost of services is used for all cases. The criteria for estimating the total capital investment required for the cases are summarized in Table 7.2-1 while Tables 7.2-3 through 7.2-8 summarize the investment requirements for each of the cases in mid 1982 dollars as follows:

Case	Description	Table
ML	Moving bed gasification Low temperature purification	7.2-3
MH	Moving bed gasification High temperature purification	7.2-4
FL	Fluidized bed gasification Low temperature purification	7.2-5
FH	Fluidized bed gasification High temperature purification	7.2-6
EL	Entrained flow gasification Low temperature purification	7.2-7
EH	Entrained flow gasification High temperature purification	7.2-8

In general for the low temperature purification cases (cases ML, FL and EL) the estimated capital costs of this study were developed using previous estimates which incorporated similar processing units as a basis for scaling and factoring. Numerous studies exist in the public domain on the economics of producing fuel gas by coal gasification and on the economics of coal gasifi-

cation combined cycle systems. The estimated costs of this study were developed using the cost estimates of similar processing facilities specifically contained in the Electric Power Research Institute Reports EPRI AF-242 and EPRI AF-244 and in the U.S. Department of Energy Report DOE/ET/14672-T1, as well as in-house estimates previously prepared by Ebastec. The overall capacity of facilities in these base estimates are all considerably greater than those required in this study but it should be noted that in all these base estimates the overall capacity was met by the use of multiple parallel facilities which reducs the degree of capacity factoring required on a train basis.

To determine the effect of high temperature purification on the capital requirements, the estimated capital costs for the conceptual high temperature sulphur and particulate removal facilities assocaited with a given gasifier were substituted for the corresponding cost of low temperature purification. Actually the adjustment was not a mere substitution of sulphur, and particulate removal costs per se, but required some adjustment of the costs of most of the other facilities including heat recovery and cooling of the raw gas, sulphur recovery, and those of the combined cycle system.

On the basis of operating parameters established by the work at MERC a conceptual process design was developed for the high temperature removal of sulphur by zinc ferrite and for regeneration of the sulphided sorbent using low concentrations of air (5% to 10%) in recycled nitrogen-sulphur dioxide gas. Major items of equipment such as pressure vessels, exchangers, and recycle compressors were sized and specified sufficiently to permit estimating their cost. The other items necessary to implement the facilities were factored on the basis of the major equipment cost. Other requirements associated with high temperature purification such as cyclones and ceramic fiber filters were developed to an extent sufficient to permit making allowances for them.

In all cases coal has been assumed delivered to the site washed and sized. If this was not true then each of the cases would require additional handling of coal and equipment for it. Such would have a larger effect on the alternate cases using the greater amount of coal but the effect would not be significant.

In all cases the net plant products are restricted to electric power, sulphur or sulphuric acid, and ammonia. Ash disposal has been incorporated as an operating expense.

Fuel, steam, and electric power are assumed available to the plant at the necessary conditions for start-up and emergency situations. The estimated costs are intended to cover a grass-roots installation. As such they are inclusive of facilities for:

- plant and instrument air
- potable and utility water
- fuel gas and nitrogen gas systems
- firewater
- flares
- effluent water treatment
- electrical substation and distribution
- buildings
- maintenance
- laboratory
- rail
- road

The contingency shown under plant investment is comprised of two parts. The first part is a 15 percent project contingency to cover the additional requirements that a more detailed design of a specific project would probably require. The second part is process contingency which is additionally applied to the less proven and less defined process technologies in an attempt to quantify this uncertainty. Most technologies during their development to commercial realization tend to require more expensive materials of construction, more complex equipment specifications, and often the need for additional as well as larger capacity equipment. The total process contingency for the individual cases is developed by applying a separate contingency to the individual process units based upon their evaluated state of development and accumulating the result. The total of these process contingencies is added to the project contingency to give the overall contingency which is shown for the individual cases on Tables 7.2-3 through 7.2-8.

Table 7.2-2 presents the process contingency factors which have been used for the process units of the various cases studied.

TABLE 7.2-1

BASIS FOR CAPITAL INVESTMENT ESTIMATES

Item	Basis or Criteria
Total Plant Investment (TPI)	Sum of the costs for the erected process, utility and support facilities including Contractor's overhead and profit, engineering and design costs, and contingencies.
Project Contingency	A project contingency factor of 15 percent is used to cover the additional equipment that might result from a more detailed design of a specific project at a defined site.
Process Contingency	This factor is applied, in addition to the project contingency, to some of the unproven and developing process technologies in an effort to quantify uncertainty in the design, performance and cost of the facilities. The allowances are given in Table 7.2-2.
Total Depreciable Capital Requirements	Includes: - Total plant investment - Royalties - Preproduction or startup costs - Construction loan interest - Initial chemical and catalyst charge - Sales taxes
Royalties	Royalty information is proprietary and, for some developing technology, not established. For comparative purposes an allowance of 2.5 percent of the gasification and purification facilities investment has been assumed to reflect paid-up royalties.
Startup Costs	Taken as 20 percent of the total annual gross operating cost plus 1.5 percent of total plant investment plus 1 month coal cost at full rate (215 MW).

TABLE 7.2-1 Contd

Item	Basis or Criteria
Construction Loan Interest	An allowance equal to the total plant investment x 1.5 years x 10% interest is used to represent the cost of the construction loan.
Sales Tax	40 percent of the total plant investment is assumed subject to sales tax of 5%. This is assumed true for Canada, although Alberta tax is 0 percent.
Working Capital	Allow 2 months for gross operating cost plus 1 month's supply of coal at full plant capacity plus 1 month's cost of ash disposal at full plant capacity.
Plant Investment Derivation of Costs	These are derived from United States costs. Based on costing experience, material and equipment are assumed equal to 60 percent of the Capital Cost. Based on supply of this material and equipment from the United States, a dollar conversion factor of 1.22 is employed. Labour Installation Cost is assumed to be same number of Canadian dollars as the American number. United States Plant Investment Costs Factor = (0.60 x 1.22) + 0.40 = 1.132.

TABLE 7.2-2

PROCESS CONTINGENCIES

Case (See subscript bottom of page)	Process Percentage					
	ML	MH	FL	FH	EL	EH
Coal Handling and Preparation	0	0	0	0	0	0
Oxygen Supply	0	0	0	0	0	0
Gasification, Ash Handling	5	5	20	20	15	15
Particulate Removal	0	20	0	20	0	20
Heat Recovery	0	NA	15	15	15	15
Gas Cooling	0	NA	0	NA	0	NA
Sulphur Removal (Selexol)	0	NA	0	NA	0	NA
Sulphur Removal (Zinc Ferrite)	NA	20	NA	20	NA	20
Sulphur Reovery and Tail Gas Cleanup	0	NA	0	NA	0	NA
Sulphuric Acid Plant	NA	0	NA	0	NA	0
Combined Cycle Facilities	5	5	5	5	5	5
Steam, Condensate and BFW Systems including Water Treatment	0	0	0	0	0	0
Cooling Water Supply	0	0	0	0	0	0
Process Condensate Treatment	0	NA	0	NA	NA	NA
Support Facilities	0	0	0	0	0	0

NOTE: NA - Not applicable
ML - Moving Bed Lurgi: - Low Temperature
MH - Moving Bed Lurgi: - High Temperature
FL - Fluidized Bed Westinghouse - Low Temperature
FH - Fluidized Bed Westinghouse - High Temperature
EL - Entrained Bed Texaco - Low Temperature
EH - Entrained Bed Texaco - High Temperature

TABLE 7.2-3

CAPITAL REQUIREMENTS - CASE ML

(Mid 1982 Dollars)

GASIFICATION: Moving Bed at Low Temperature (Lurgi)
PURIFICATION: Low Temperature Sulphur Removal (Selexol)

Plant Investment	$1000
Coal Handling	10,784
Oxygen Supply including air separation and compression of air and oxygen	29,700
Gasification including ash handling, particulate removal, heat recovery and gas cooling	44,721
Purification including sulphur removal, sulphur recovery and tail gas treatment	24,227
Combined Cycle Facilities including combustion gas turbines, heat recovery steam generation, gas and steam turbine driven power generation	92,079
Process Condensate Treatment including tar oil separation, phenol extration, and ammonia recovery	53,377
Steam, Condensate and Boiler Feedwater System including water treatment	6,615
Cooling Water Supply including cooling towers	6,954
Other Utility and Support Facilities	37,477
Subtotal	305,914
Contingency	52,014
Total Plant Investment (TPI)	357,929
Capital Charges	
Paid-up Royalties	1,724
Start-up Costs	17,193
Construction Loan Interest	53,689
Initial Chemical and Catalyst Charges	3,760
Sales Taxes	7,159
Total Capital Charges	83,525
Depreciable Capital	441,450
Working Capital	10,828
Total Capital Required	452,278
Plant Investment for 215 KW - $/KW	1,665
Total Capital for 215 KW - $/KW	2,104

TABLE 7.2-4

CAPITAL REQUIREMENTS - CASE MH

(Mid 1982 Dollars)

GASIFICATION: Moving Bed at Low Temperature (Lurgi)
PURIFICATION: High Temperature Purification

Plant Investment	$1000
Coal Handling	9,440
Oxygen Supply including air separation and compression of air and oxygen	24,931
Gasification including ash handling	27,453
Particulate Removal	16,121
Sulphur Removal and Regeneration	31,141
Sulphuric Acid	5,529
Combined Cycle Facilities including combustion gas turbines, heat recovery steam generation, gas and steam turbine driven power generation	92,931
Steam, Condensate and Boiler Feedwater System including water treatment	5,220
Cooling Water Supply including cooling towers	5,254
Other Utility and Support Facilities	33,795
Subtotal	251,815
Contingency	53,244
Total Plant Investment (TPI)	305,059
Capital Charges	
Paid-up Royalties	1,868
Start-up Costs	17,835
Construction Loan Interest	45,759
Initial Chemical and Catalyst Charges	3,332
Sales Taxes	6,100
Total Capital Charges	75,895
Depreciable Capital	380,954
Working Capital	11,532
Total Capital Required	392,486
Plant Investment for 215 KW - $/KW	1,419
Total Capital for 215 KW - $/KW	1,826

TABLE 7.2-5

CAPITAL REQUIREMENTS - CASE FL

(Mid 1982 Dollars)

GASIFICATION: Fluidized Bed at Moderate Temperature (Westinghouse)
PURIFICATION: Low Temperature Sulphur Removal (Selexol)

Plant Investment	$1000
Coal Handling including grinding, drying and feeding	18,424
Oxygen Supply including air separation and compression of air and oxygen	33,130
Gasification including ash handling, heat recovery, particulate removal and gas cooling	31,664
Purification including sulphur removal, sulphur recovery and tail gas treatment	15,941
Combined Cycle Facilities including combustion gas turbines, heat recovery steam generation, gas and steam turbine driven power generation	98,000
Process Condensate Treatment including ammonia recovery	1,465
Steam, Condensate and Boiler Feedwater System including water treatment	3,322
Cooling Water Supply including cooling towers	6,432
Other Utility and Support Facilities	35,202
Subtotal	243,580
Contingency	46,426
Total Plant Investment (TPI)	290,006
Capital Charges	
Paid-up Royalties	1,190
Start-up Costs	14,470
Construction Loan Interest	43,500
Initial Chemical and Catalyst Charges	1,570
Sales Taxes	5,800
Total Capital Charges	66,530
Depreciable Capital	356,536
Working Capital	9,261
Total Capital Required	365,797
Plant Investment for 215 KW - $/KW	1,349
Total Capital for 215 KW - $/KW	1,701

TABLE 7.2-6

CAPITAL REQUIREMENTS - CASE FH

(Mid 1982 Dollars)

GASIFICATION: Fluidized Bed at Moderate Temperature (Westinghouse)
PURIFICATION: High Temperature Purification

Plant Investment	$1000
Coal Handling including grinding, drying and feeding	17,812
Oxygen Supply including air separation and compression of air and oxygen	31,670
Gasification including ash handling	24,490
Particulate Removal	10,468
Sulphur Removal and Regeneration	27,813
Sulphuric Acid	5,692
Combined Cycle Facilities including combustion gas turbines, heat recovery steam generation, gas and steam turbine driven power generation	99,142
Process Condensate Treatment	368
Steam, Condensate and Boiler Feedwater System including water treatment	2,887
Cooling Water Supply including cooling towers	6,546
Other Utility and Support Facilities	34,237
Subtotal	261,125
Contingency	61,463
Total Plant Investment (TPI)	322,588
Capital Charges	
Paid-up Royalties	1,569
Start-up Costs	18,362
Construction Loan Interest	48,388
Initial Chemical and Catalyst Charges	2,284
Sales Taxes	6,452
Total Capital Charges	77,055
Depreciable Capital	399,643
Working Capital	11,776
Total Capital Required	411,419
Plant Investment for 215 KW - $/KW	1,500
Total Capital for 215 KW - $/KW	1,914

TABLE 7.2-7

CAPITAL REQUIREMENTS - CASE EL

(Mid 1982 Dollars)

GASIFICATION: Entrained Bed at High Temperature (Texaco)
PURIFICATION: Low Temperature Sulphur Removal (Selexol)

Plant Investment	$1000
Coal Handling including grinding and slurry preparation	26,378
Oxygen Supply including air separation and compression of air and oxygen	46,244
Gasification including ash handling, heat recovery, particulate removal, and gas cooling	54,842
Purification including sulphur removal, sulphur recovery and tail gas treatment	15,330
Combined Cycle Facilities including combustion gas turbines, heat recovery steam generation, gas and steam turbine driven power generation	100,411
Steam, Condensate and Boiler Feedwater System including water treatment	2,564
Cooling Water Supply including cooling towers	6,764
Other Utility and Support Facilities	34,636
Subtotal	287,169
Contingency	54,576
Total Plant Investment (TPI)	341,745
Capital Charges	
Paid-up Royalties	1,755
Start-up Costs	15,754
Construction Loan Interest	51,261
Initial Chemical and Catalyst Charges	1,253
Sales Taxes	6,835
Total Capital Charges	76,858
Depreciable Capital	418,603
Working Capital	9,753
Total Capital Required	428,356
Plant Investment for 215 KW - $/KW	1,590
Total Capital for 215 KW - $/KW	1,992

TABLE 7.2-8

CAPITAL REQUIREMENTS - CASE EH

(Mid 1982 Dollars)

GASIFICATION: Entrained Bed at High Temperature (Texaco)
PURIFICATION: High Temperature Purification

Plant Investment	$1000
Coal Handling including grinding and slurry preparation	25,249
Oxygen Supply including air separation and compression of air and oxygen	43,625
Gasification including ash handling	39,371
Particulate Removal	10,797
Sulphur Removal and Regneration	32,092
Sulphuric Acid	5,581
Combined Cycle Facilities including combustion gas turbines, heat recovery steam generation, gas and steam turbine driven power generation	101,444
Steam, Condensate and Boiler Feedwater System including water treatment	1,040
Cooling Water Supply including cooling towers	7,039
Other Utility and Support Facilities	33,419
Subtotal	299,657
Contingency	64,505
Total Plant Investment (TPI)	364,162
Capital Charges	
Paid-up Royalties	2,057
Start-up Costs	19,249
Construction Loan Interest	54,624
Initial Chemical and Catalyst Charges	2,136
Sales Taxes	7,282
Total Capital Charges	85,348
Depreciable Capital	449,510
Working Capital	11,979
Total Capital Required	461,489
Plant Investment for 215 KW - $/KW	1,694
Total Capital for 215 KW - $/KW	2,146

7.3 Operating Requirements

A consistent basis for estimating the annual operating cost, which is outlined in Table 7.3-1, is used for all cases. Tables 7.3-2 through 7.3-7 present for each case the operating requirements per megawatt of net electric power produced as well as the annual operating cost of the plants on the basis of a 70 percent operating factor. Monetary values are expressed in mid-1982 dollars.

Case	Description	Table
ML	Moving bed gasification Low temperature purification	7.3-2
MH	Moving bed gasification High temperature purification	7.3-3
FL	Fluidized bed gasification Low temperature purification	7.3-4
FH	Fluidized bed gasification High temperature purification	7.3-5
EL	Entrained flow gasification Low temperature purification	7.3-6
EH	Entrained flow gasification High temperature purification	7.3-7

TABLE 7.3-1

BASIS FOR ESTIMATING ANNUAL OPERATING COST

Item	Basis or Criteria
(Monetary values expressed in mid-1982 dollars)	
Operating Load Factor	70 percent
Cost of Coal Delivered	$2.00 per million kilo joules
Raw Water	$0.015 per cubic meter
Ash Disposal	$2.00 per metric tonme (1000 kg)
By-product Ammonia Credit	$110. per metric tonme
By-product Sulphur Credit	$ 30. per metric tonne
By-product Sulphuric Acid credit	$ 30. per metric tonne
Maintenance	Annual maintenance costs (labour and materialsd allowed for at 2.5 percent of the total plant investment. The total cost is distributed 60 percent to labour and 40 percent to materials.
Operating Labour	At an average rate of $15 per manhour, inclusive of fringe benefits and allowing for vacations, holidays, and swing shifts.
Maintenance Labour	60 percent of total maintenance cost.
Administrative and Supervisory Labour	30 percent of operating and maintenance labour.
General and Administrative Expense	2 percent of total plant investment per year.
Supplies	Operating supplies at 30 percent of operating labour plus maintenance materials at 40 percent of total maintenance.
Property Taxes and Insurance	2½ percent of total plant investment per year.
Basis for Estimating Annual Capital Charges	
Depreciation Expense	Based on total depreciable capital, ammortized over a period of 25 years (assumed plant life) on a straight line basis.
Average Return on Capital	Based on an assumed required rate of return on Total Capital Required (debt plus equity) of 12.5 percent per year.
Average Tax	Tax on business profits assumed to be zero (i.e the public utility case).

TABLE 7.3-2

OPERATING REQUIREMENTS AND COSTS - CASE ML

(Mid 1982 Dollars)

GASIFICATION: Moving Bed at Low Temperature (Lurgi)
PURIFICATION: Low Temperature Sulphur Removal (Selexol)

Operating Requirements

Coal required (HHV)	-	11.262×10^6 kJ/MW
Raw Water required	-	4.55 m^3/MW
Catalyst and Chemicals required	-	$0.390/MW
Operating Personnel required	-	12 persons per 8-hour shift
Sulphur produced	-	13.17 kg/MW
Ammonia produced	-	4.90 kg/MW
Ash produced for disposal	-	38.0 kg/MW

Estimated Annual Operating Cost

Net Power at 100% Capacity	-	215 MW-h/h
Operating Factor	-	70 percent
Net Annual Power	-	1,318,380 MW-h

	$1000
Coal required	29,695
Raw Water required	90
Catalysts and Chemicals	514
Operating Labour (105,120 manhours x $15)	1,577
Maintenance Labour	4,743
Supervision	1,896
Operating and Maintenance Materials	3,635
General and Administrative Expense	6,324
Property Taxes and Insurance Expense	7,905
Ash Disposal Expense	100
Total Gross Operating Cost Per Year	56,479
Credit for Sulphur Produced	(517)
Credit for Ammonia Produced	(711)
Total Credits for By-products Per Year	(1,228)
Total Net Operating Cost Per Year	55,251

Capital Charges

Depreciation	17,658
Average Return on Capital (12.5%)	56,535
Average Tax	-
Total Capital Charges per year	74,193
Total Cost of Service per year	129,444
Cost of power, $/KW-h	0.0982

TABLE 7.3-3

OPERATING REQUIREMENTS AND COSTS - CASE MH

(Mid 1982 Dollars)

GASIFICATION: Moving Bed at Low Temperature (Lurgi)
PURIFICATION: High Temperature Purification

Operating Requirements

Coal required (HHV)	- 9.049×10^6 kJ/MW
Raw Water required	- 2.6 m^3/MW
Catalyst and Chemicals required	- $4.49/MW
Operating Personnel required	- 10 persons per 8-hour shift
Sulphuric Acid produced	- 35.29 kg/MW
Ash produced for disposal	- 30.5 kg/MW

Estimated Annual Operating Cost

Net Power at 100% Capacity	- 215 MW-h/h
Operating Factor	- 70 percent
Net Annual Power	- 1,318,380 MW-h

	$1000
Coal required	23,860
Raw Water required	54
Catalysts and Chemicals	5,920
Operating Labour (87,600 manhours x $15)	1,314
Maintenance Labour	4,042
Supervision	1,607
Operating and Maintenance Materials	3,089
General and Administrative Expense	5,390
Property Taxes and Insurance Expense	6,737
Ash Disposal Expense	80
Total Gross Operating Cost Per Year	52,093
Credit for Sulphuric Acid Produced	(1,396)
Total Net Operating Cost Per Year	50,697

Capital Charges

Depreciation	15,238
Average Return on Capital (12.5%)	49,061
Average Tax	-
Total Capital Charges per year	64,299
Total Cost of Service per year	114,996
Cost of power, $/KW-h	0.0872

TABLE 7.3-4

OPERATING REQUIREMENTS AND COSTS - CASE FL

(Mid 1982 Dollars)

GASIFICATION: Fluidized Bed at Moderate Temperature (Westinghouse)
PURIFICATION: Low Temperature Sulphur Removal (Selexol)

Operating Requirements

Coal required (HHV)	-	10.047×10^6 kJ/MW
Raw Water required	-	3.70 m^3/MW
Catalyst and Chemicals required	-	$0.264/MW
Operating Personnel required	-	9 persons per 8-hour shift
Sulphur produced	-	11.75 kg/MW
Ammonia produced	-	0.59 kg/MW
Ash produced for disposal	-	33.9 kg/MW

Estimated Annual Operating Cost

Net Power at 100% Capacity	-	215 MW-h/h
Operating Factor	-	70 percent
Net Annual Power	-	1,318,380 MW-h

	$1000
Coal required	26,492
Raw Water required	72
Catalysts and Chemicals	348
Operating Labour (78,840 manhours x $15)	1,183
Maintenance Labour	3,843
Supervision	1,508
Operating and Maintenance Materials	2,917
General and Administrative Expense	5,124
Property Taxes and Insurance Expense	6,405
Ash Disposal Expense	89
Total Gross Operating Cost Per Year	47,981
Credit for Sulphur Produced	(465)
Credit for Ammonia Produced	(86)
Total Credits for By-products Per Year	(551)
Total Net Operating Cost Per Year	47,430
Capital Charges	
Depreciation	14,261
Average Return on Capital (12.5%)	45,725
Average Tax	-
Total Capital Charges per year	60,086
Total Cost of Service per year	107,516
Cost of power, $/KW-h	0.0816

TABLE 7.3-5

OPERATING REQUIREMENTS AND COSTS - CASE FH

(Mid 1982 Dollars)

GASIFICATION: Fluidized Bed at Moderate Temperature (Westinghouse)
PURIFICATION: High Temperature Purification

Operating Requirements

Coal required (HHV)	-	9.497×10^6 kJ/MW
Raw Water required	-	2.8 m^3/MW
Catalyst and Chemicals required	-	$3.185/MW
Operating Personnel required	-	10 persons per 8-hour shift
Sulphuric Acid produced	-	37.04 kg/MW
Ash produced for disposal	-	32.9 kg/MW

Estimated Annual Operating Cost

Net Power at 100% Capacity	-	215 MW-h/h
Operating Factor	-	70 percent
Net Annual Power	-	1,318,380 MW-h

	$1000
Coal required	25,041
Raw Water required	54
Catalysts and Chemicals	4,199
Operating Labour (87,600 manhours x $15)	1,314
Maintenance Labour	4,275
Supervision	1,677
Operating and Maintenance Materials	3,244
General and Administrative Expense	5,699
Property Taxes and Insurance Expense	7,124
Ash Disposal Expense	84
Total Gross Operating Cost Per Year	52,711
Credit for Sulphuric Acid Produced	1,465)
Total Net Operating Cost Per Year	51,245

Capital Charges

Depreciation	15,986
Average Return on Capital (12.5%)	51,427
Average Tax	-
Total Capital Charges per year	67,413
Total Cost of Service per year	118,658
Cost of power, $/KW-h	0.0900

TABLE 7.3-6

OPERATING REQUIREMENTS AND COSTS - CASE EL

(Mid 1982 Dollars)

GASIFICATION: Entrained Bed at High Temperature (Texaco)
PURIFICATION: Low Temperature Sulphur Removal (Selexol)

Operating Requirements

Coal required (HHV)	-	9.885×10^6 kJ/MW
Raw Water required	-	2.91 m^3/MW
Catalyst and Chemicals required	-	$0.344/MW
Operating Personnel required	-	9 persons per 8-hour shift
Sulphur produced	-	11.56 kg/MW
Ammonia produced	-	Zero
Ash produced for disposal	-	33.3 kg/MW

Estimated Annual Operating Cost

Net Power at 100% Capacity	-	215 MW-h/h
Operating Factor	-	70 percent
Net Annual Power	-	1,318,380 MW-h

	$1000
Coal required	26,064
Raw Water required	57
Catalysts and Chemicals	454
Operating Labour (78840 manhours x $15)	1,183
Maintenance Labour	4,528
Supervision	1,713
Operating and Maintenance Materials	3,374
General and Administrative Expense	6,038
Property Taxes and Insurance Expense	7,547
Ash Disposal Expense	88
Total Gross Operating Cost Per Year	51,046
Credit for Sulphur Produced	(457)
Credit for Ammonia Produced	-
Total Credits for By-products Per Year	(457)
Total Net Operating Cost Per Year	50,589

Capital Charges

Depreciation	16,794
Average Return on Capital (12.5%)	53,545
Average Tax	-
Total Capital Charges per year	70,289
Total Cost of Service per year	120,878
Cost of power, $/KW-h	0.0917

TABLE 7.3-7

OPERATING REQUIREMENTS AND COSTS - CASE EH

(Mid 1982 Dollars)

GASIFICATION: Entrained Bed at High Temperature (Texaco)
PURIFICATION: High Temperature Sulphur Removal

Operating Requirements

Coal required (HHV)	-	9.191×10^6 kJ/MW
Raw Water required	-	2.9 m^3/MW
Catalyst and Chemicals required	-	$3.081/MW
Operating Personnel required	-	10 persons per 8-hour shift
Sulphuric Acid produced	-	35.85 kg/MW
Ash produced for disposal	-	31.0 kg/MW

Estimated Annual Operating Cost

Net Power at 100% Capacity	-	215 MW-h/h
Operating Factor	-	70 percent
Net Annual Power	-	1,318,380 MW-h

	$1000
Coal required	24,235
Raw Water required	57
Catalysts and Chemicals	4,061
Operating Labour (87,600 manhours x $15)	1,314
Maintenance Labour	4,825
Supervision	1,842
Operating and Maintenance Materials	3,611
General and Administrative Expense	6,434
Property Taxes and Insurance Expense	8,042
Ash Disposal Expense	82
Total Gross Operating Cost Per Year	54,503
Credit for Sulphuric Acid Produced	(1,418)
Total Net Operating Cost Per Year	53,085

Capital Charges

Depreciation	17,980
Average Return on Capital (12.5%)	57,686
Average Tax	-
Total Capital Charges per year	75,666
Total Cost of Service per year	128,751
Cost of power, $/KW-h	0.0977

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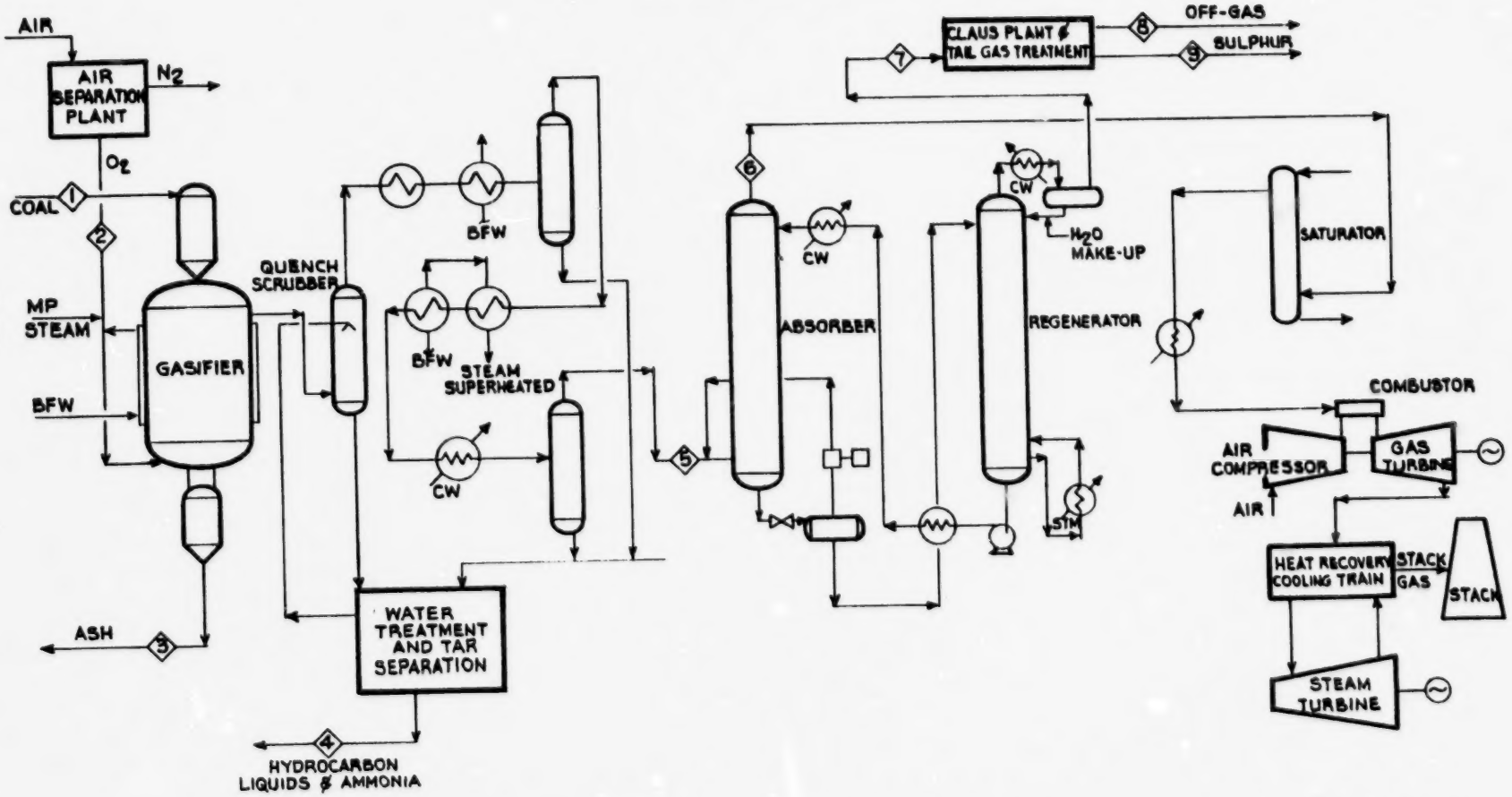
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT LURGI DRY ASH OXYGEN BLOWN GASIFIER - SELOXOL PURIFICATION - COMBINED CYCLE PLANT

FIG. 7·1-1 CASE ML



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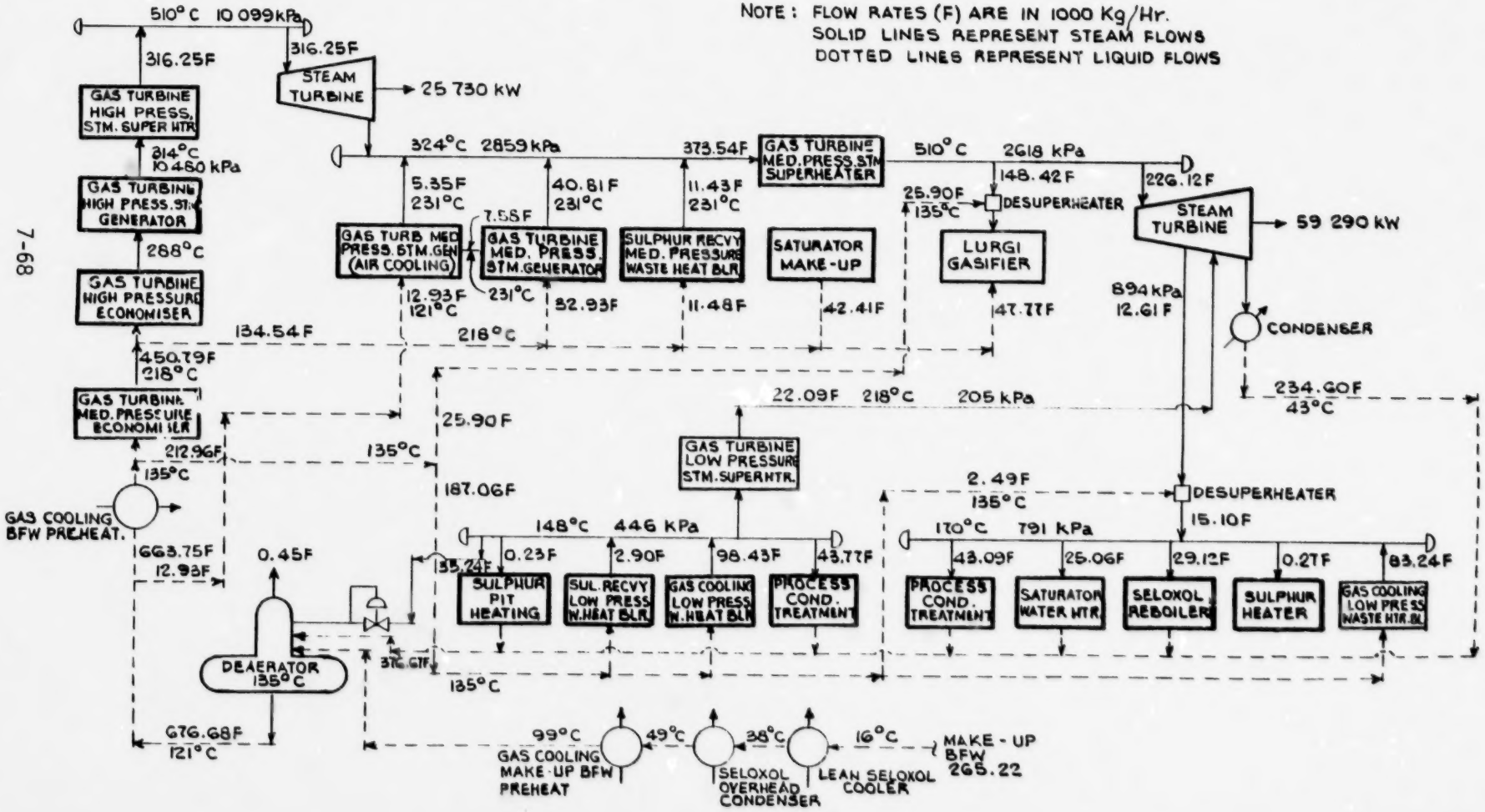
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT LURGI DRY ASH OXYGEN BLOWN GASIFIER - SELOXOL PURIFICATION - (LOW TEMPERATURE) - COMBINED CYCLE - STEAM BALANCE

FIG. 7·1-2 CASE ML



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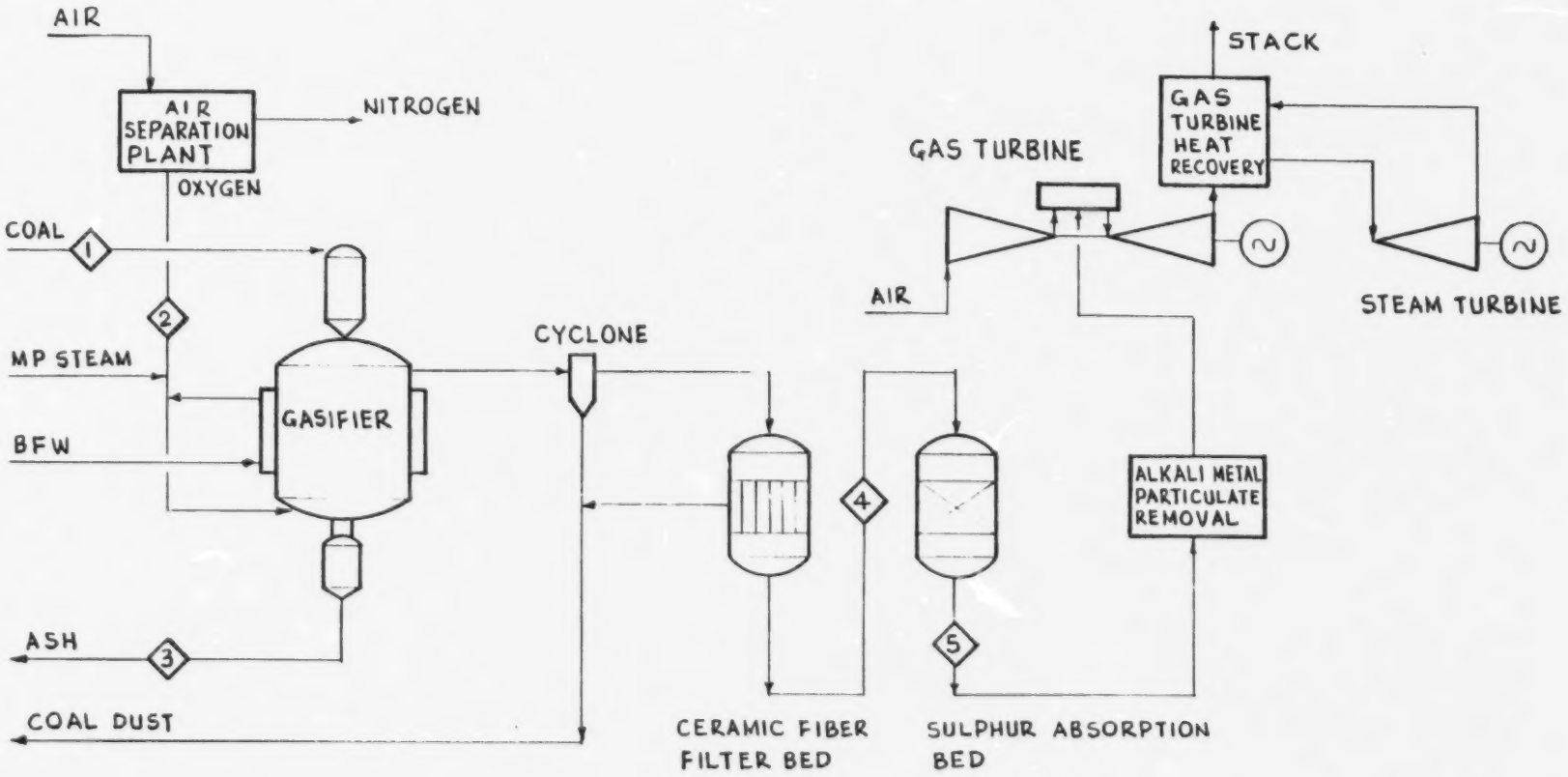
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT LURGI DRY ASH OXYGEN BLOWN GASIFIER - HIGH TEMPERATURE SULPHUR AND PARTICULATE REMOVAL FOR COMBINED CYCLE PLANT

FIG. 7·1-3 CASE MH



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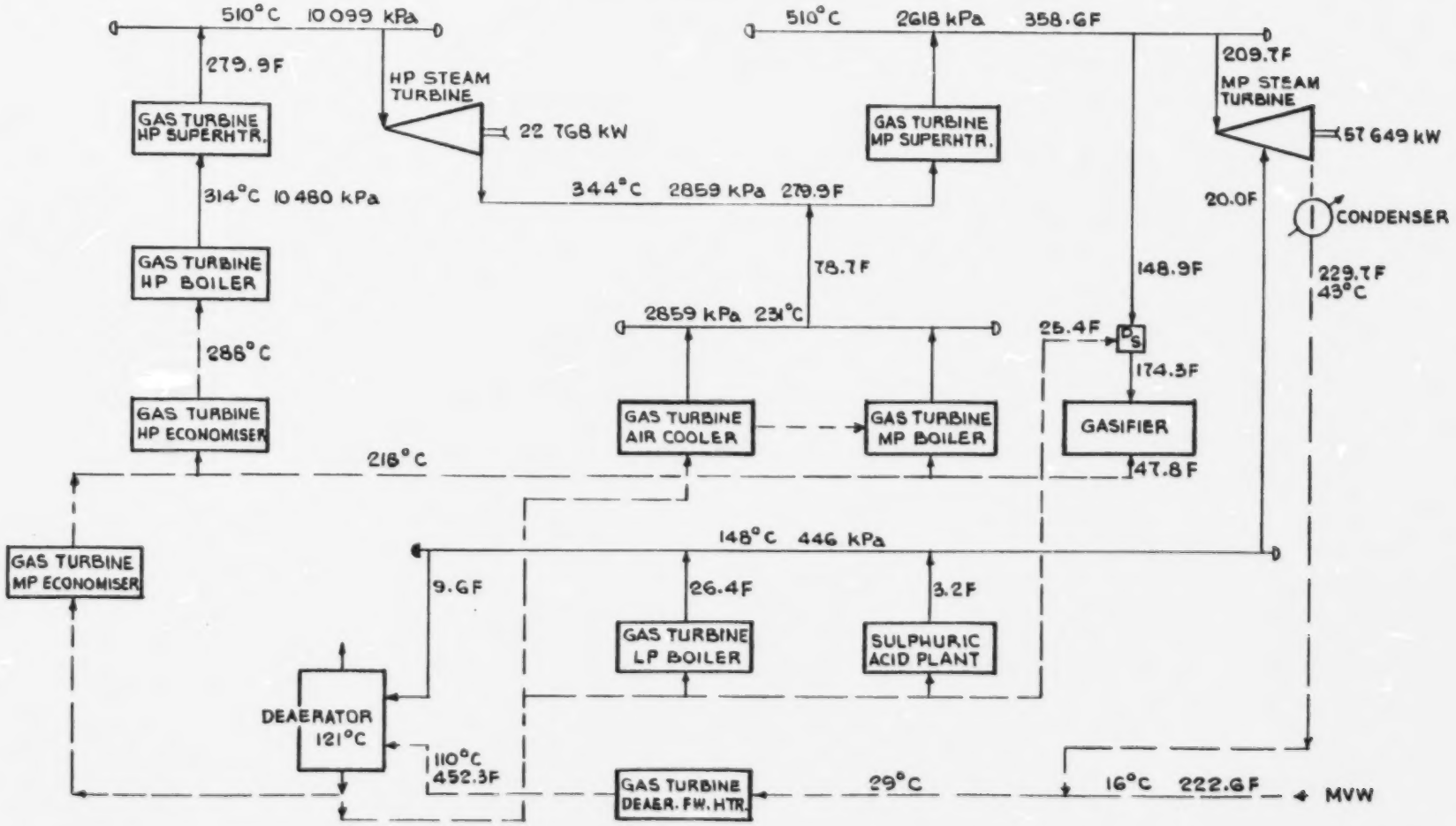
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT LURGI DRY ASH OXYGEN BLOWN GASIFIER - HIGH TEMPERATURE PURIFICATION - COMBINED CYCLE - STEAM SYSTEM

FIG. 7·1-4 CASE MH



NOTE: FLOW RATES (F) ARE IN 1000 Kg/Hr.

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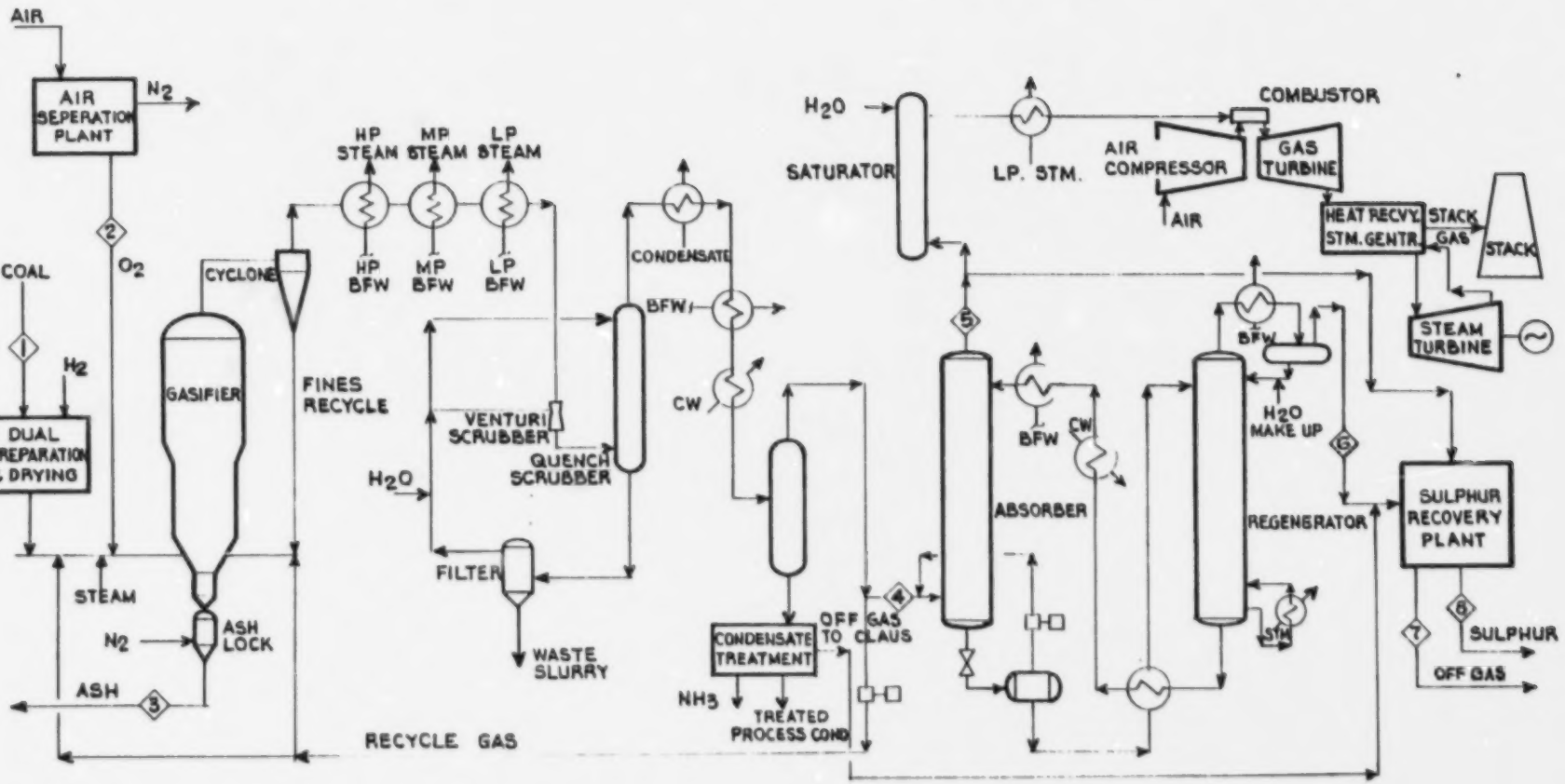
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT WESTINGHOUSE OXYGEN BLOWN GASIFIER - SELOXOL PURIFICATION - COMBINED CYCLE PLANT

FIG. 7-1-5 CASE FL



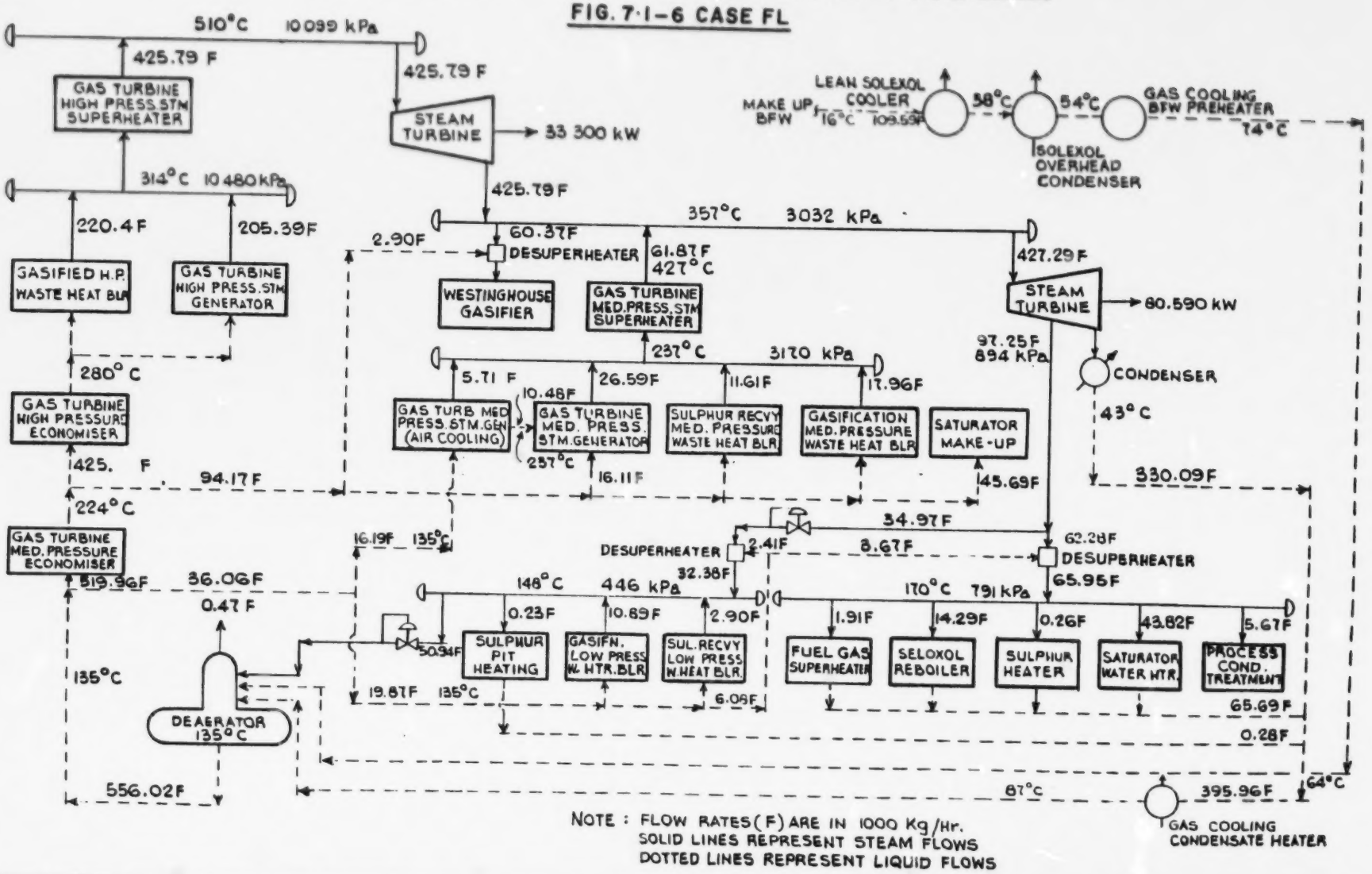
GASIFICATION | HEAT RECOVERY AND PARTICULATE REMOVAL | GAS COOLING | SELOXOL SULPHUR REMOVAL | COMBINED CYCLE

EBASTEC Lavalin Inc.

PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT WESTINGHOUSE OXYGEN BLOWN GASIFIER - LOW TEMPERATURE PURIFICATION - STEAM BALANCE

FIG. 7·1-6 CASE FL



NOTE : FLOW RATES (F) ARE IN 1000 Kg/Hr.
SOLID LINES REPRESENT STEAM FLOWS
DOTTED LINES REPRESENT LIQUID FLOWS

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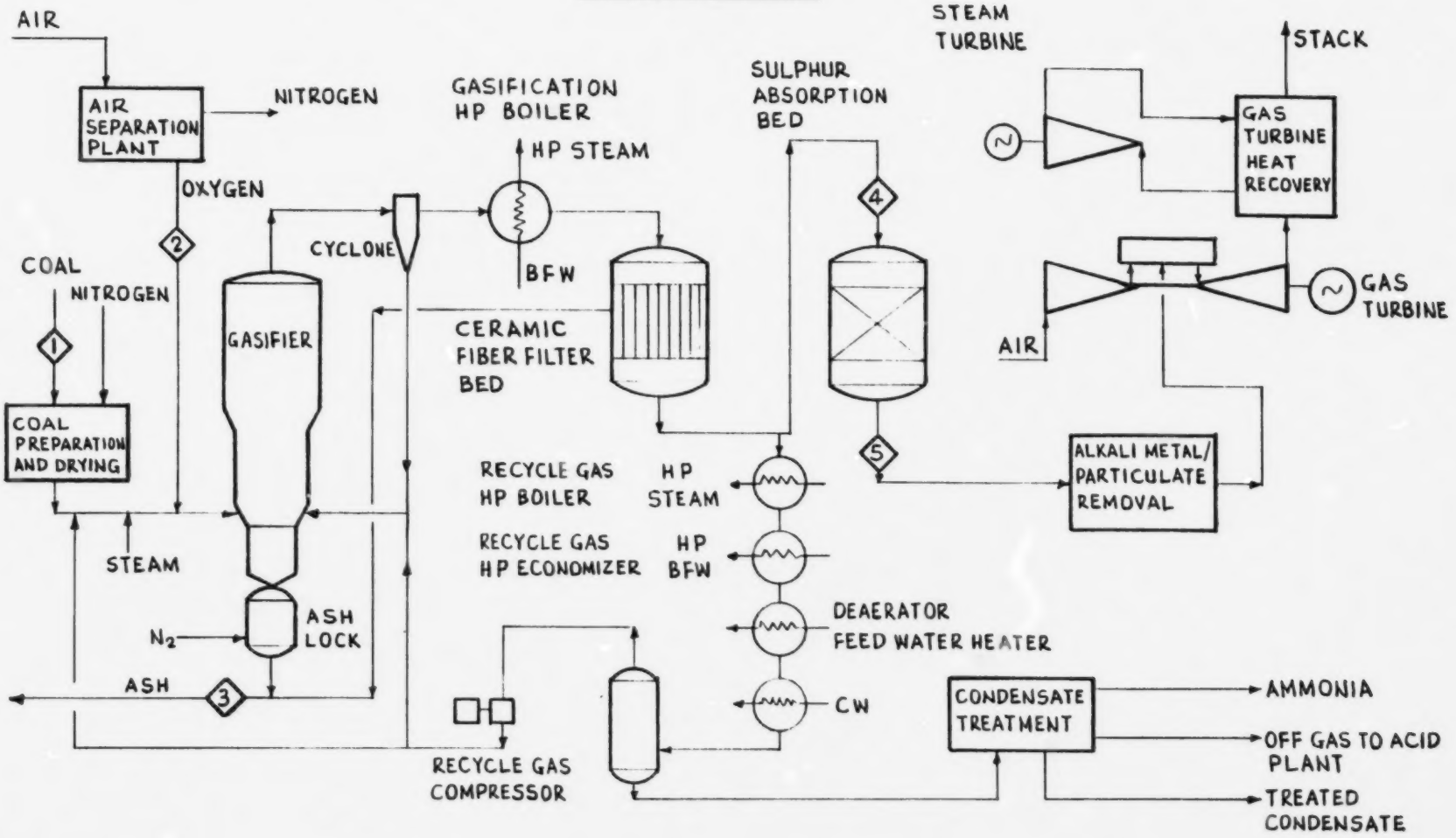
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT WESTINGHOUSE OXYGEN BLOWN GASIFIER – HIGH TEMPERATURE SULPHUR AND PARTICULATE REMOVAL FOR COMBINED CYCLE PLANT

FIG. 7·1-7 CASE FH



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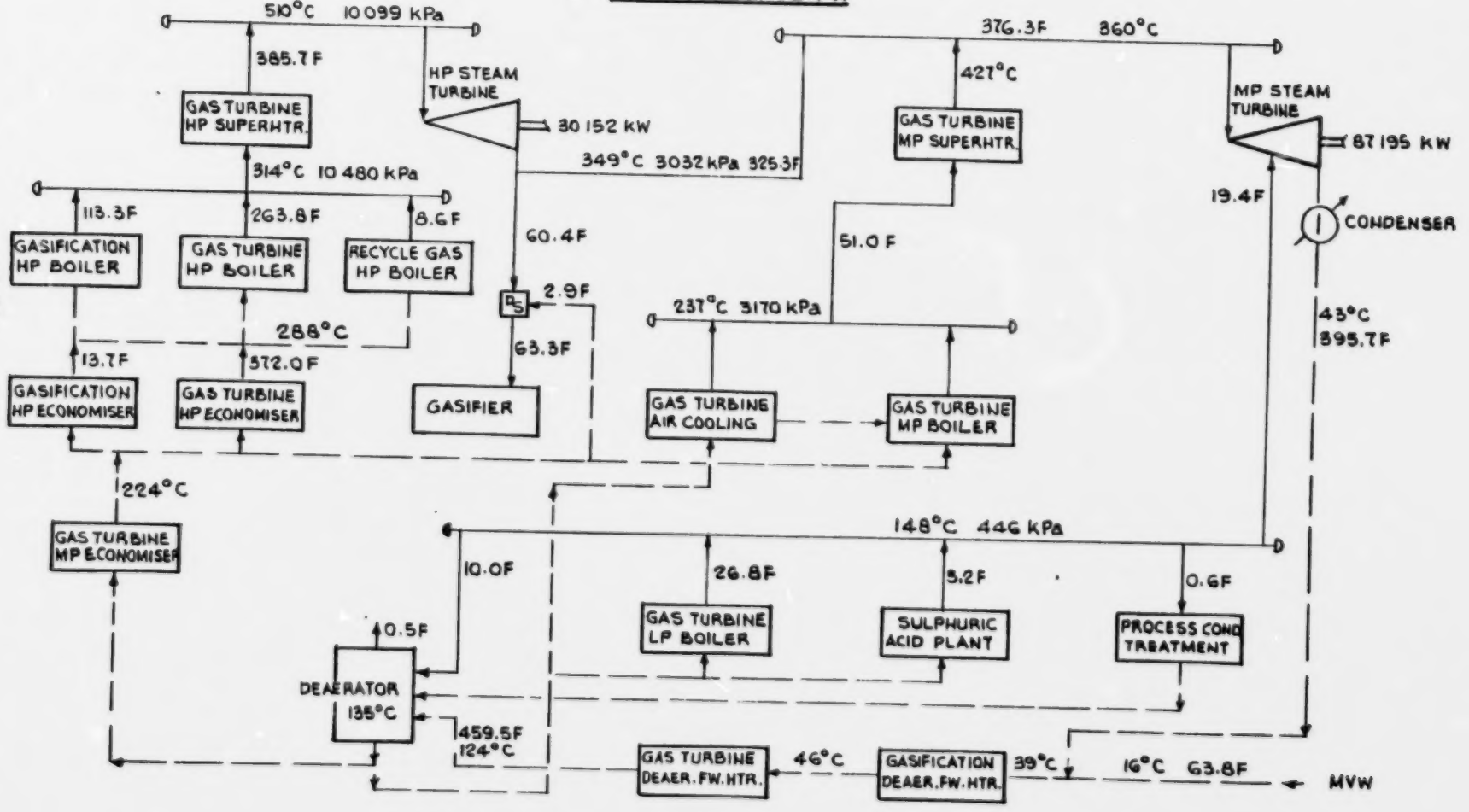
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT WESTINGHOUSE OXYGEN BLOWN GASIFIER – HIGH TEMPERATURE PURIFICATION – COMBINED CYCLE – STEAM SYSTEM

FIG.7-1-8 CASE FH



NOTE : FLOW RATES (F) ARE IN 1 000 Kg/Hr.

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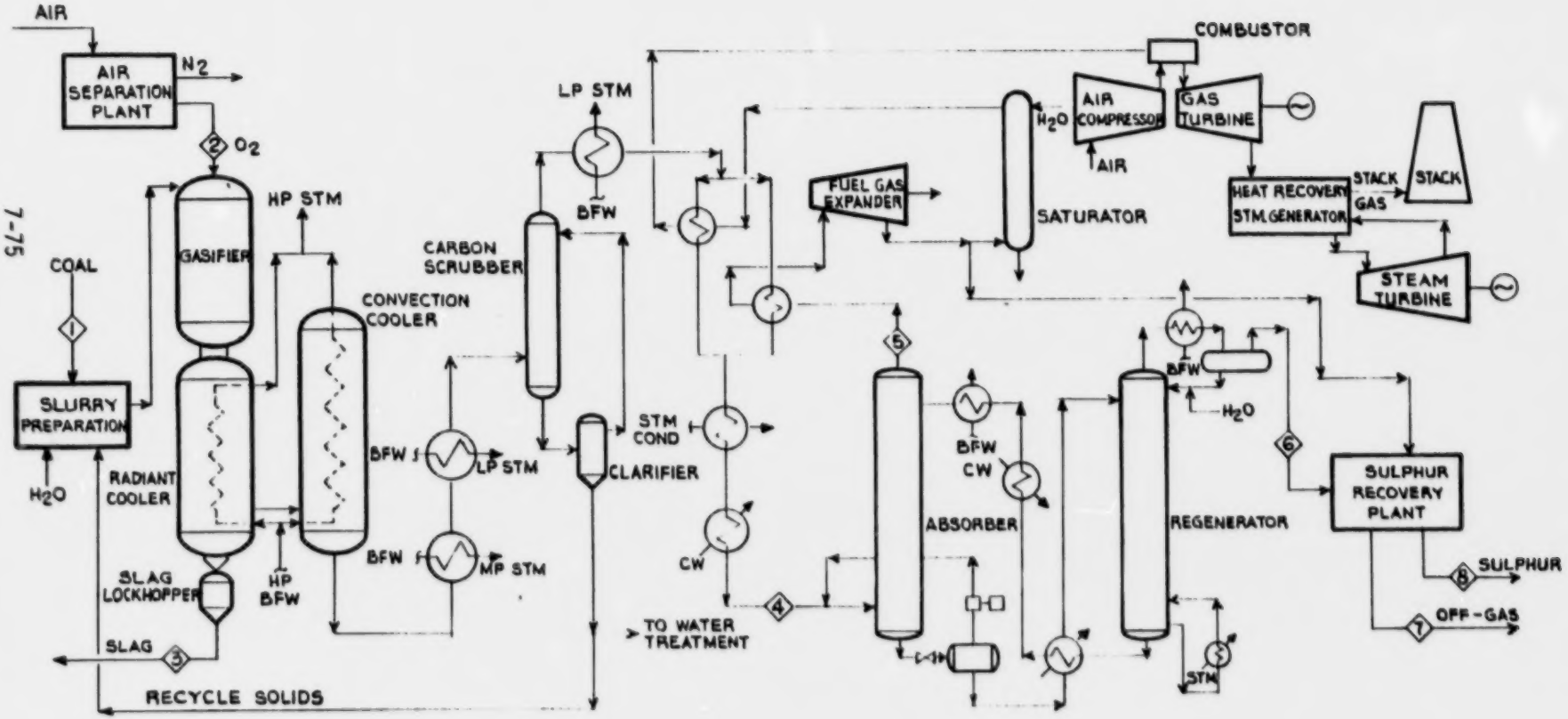
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT TEXACO OXYGEN BLOWN GASIFIER - SELOXOL PURIFICATION - COMBINED CYCLE PLANT

FIG. 7·1-9 CASE EL



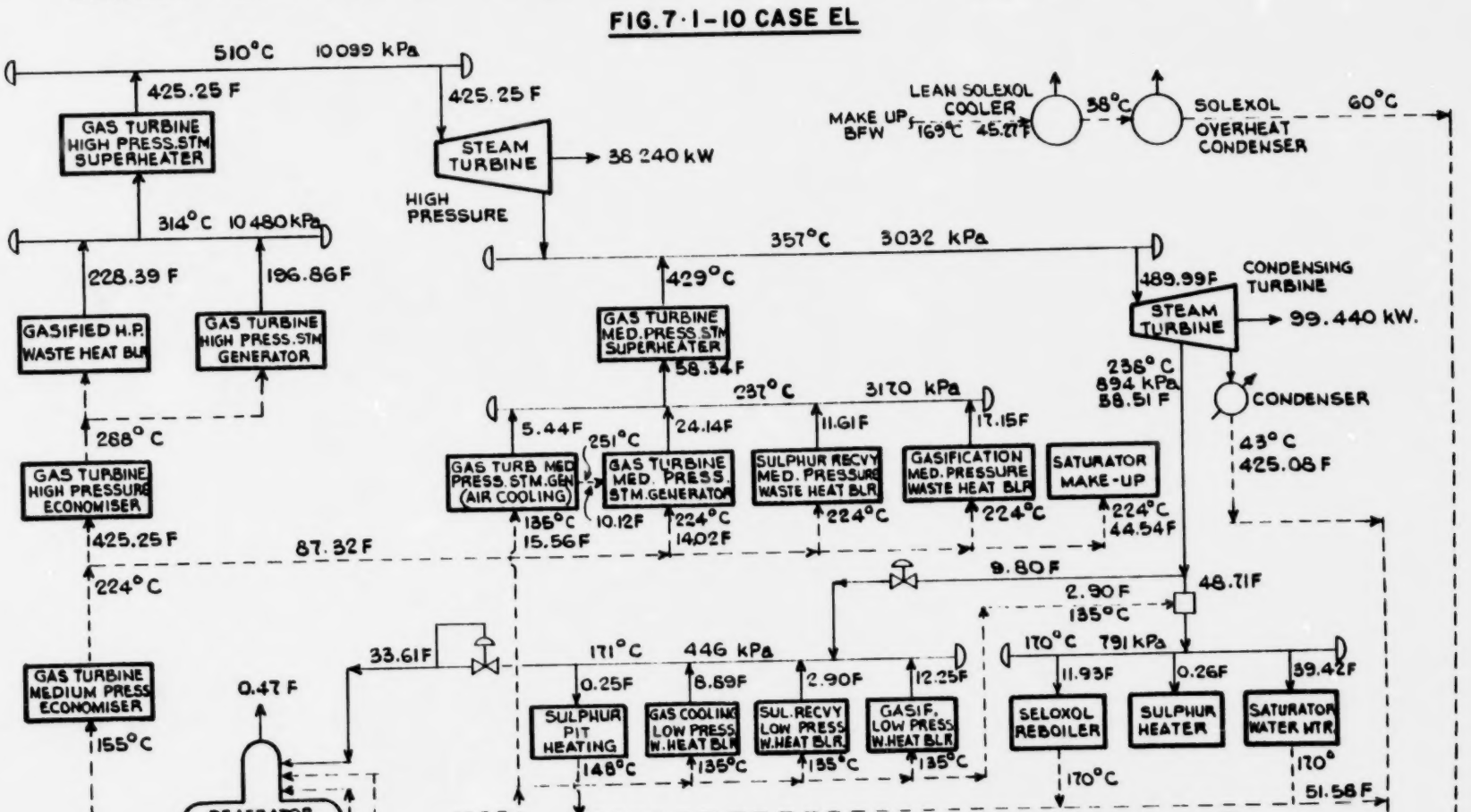
GASIFICATION | HEAT RECOVERY AND PARTICULATE REMOVAL | GAS COOLING | SELOXOL SULPHUR REMOVAL | COMBINED CYCLE

EBASTEC Lavalin Inc.

PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT TEXACO OXYGEN BLOWN GASIFIER - SELOXOL PURIFICATION - COMBINED CYCLE - STEAM BALANCE

FIG. 7-1-10 CASE EL



NOTE: FLOW RATES (F) ARE IN 1000 Kg/Hr
SOLID LINES REPRESENT STEAM FLOWS
DOTTED LINES REPRESENT LIQUID FLOWS

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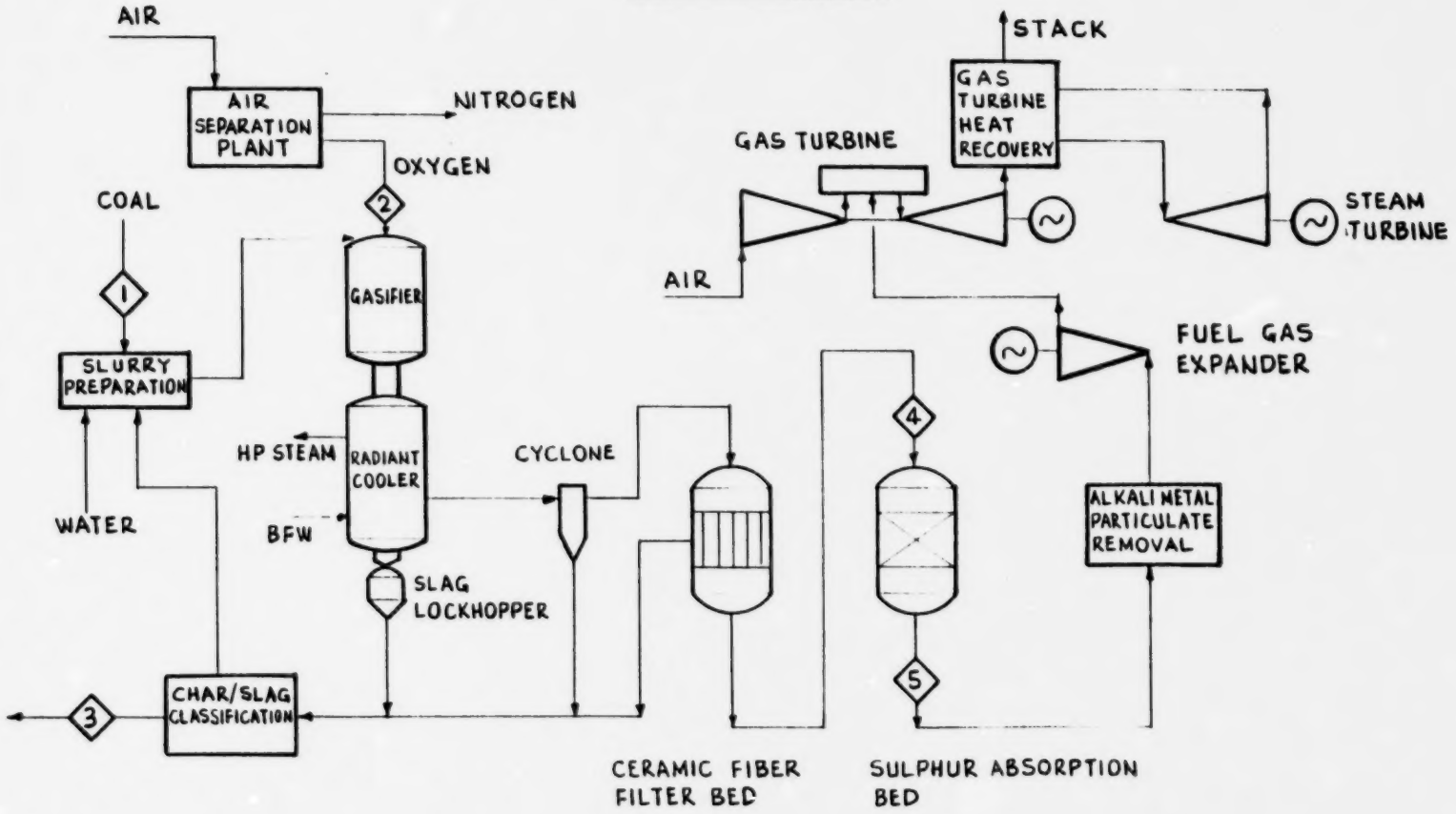
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT TEXACO OXYGEN BLOWN GASIFIER - HIGH TEMPERATURE SULPHUR AND PARTICULATE REMOVAL FOR COMBINED CYCLE PLANT

FIG.7·1-11 CASE EH



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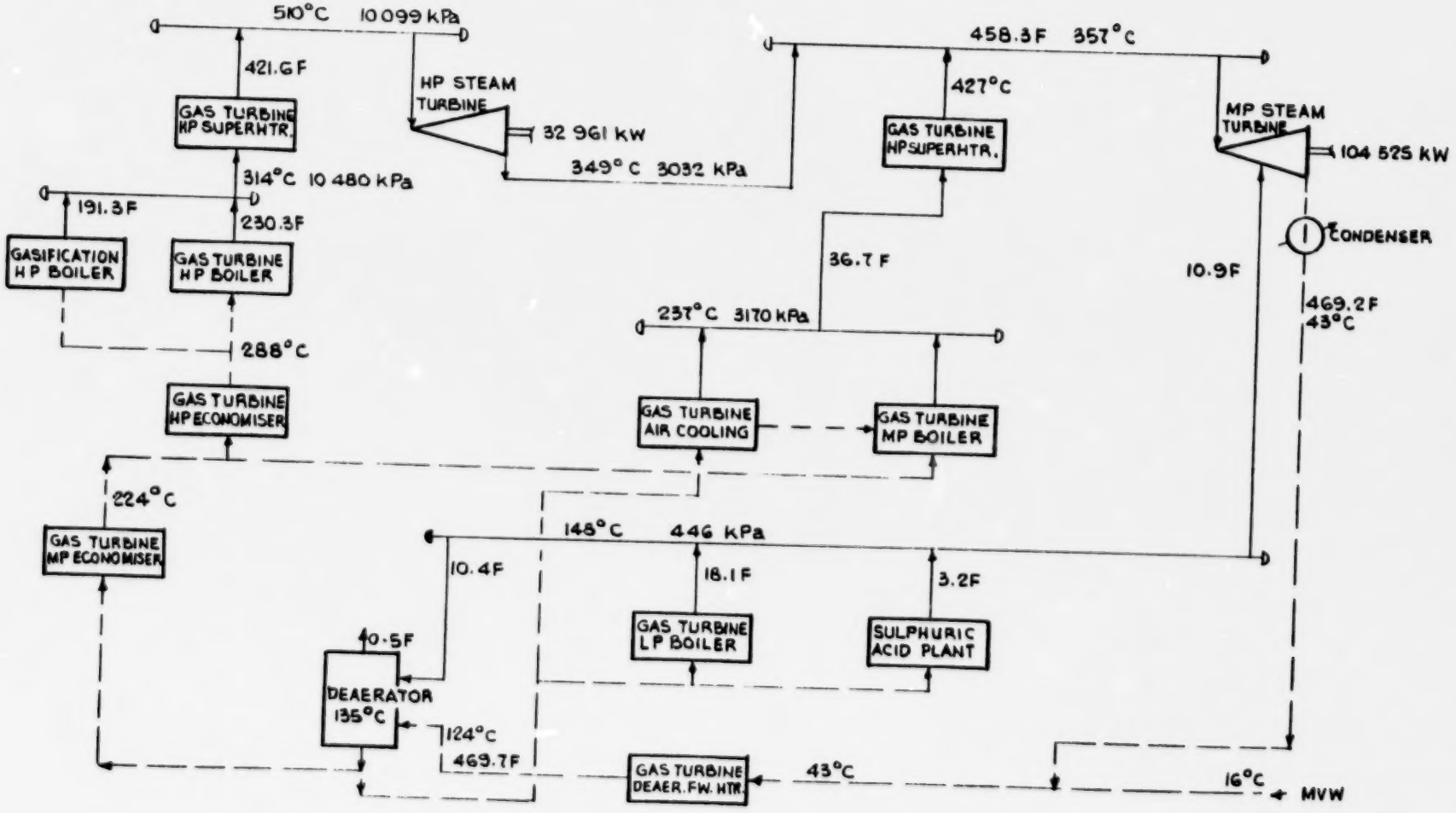
PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT TEXACO OXYGEN BLOWN GASIFIER - HIGH TEMPERATURE PURIFICATION - COMBINED CYCLE - STEAM SYSTEM

FIG. 7·1-12 CASE EH



NOTE : FLOW RATES (F) ARE IN 1000 Kg/Hr.

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PROJECT HIGH TEMPERATURE GAS CLEAN UP STUDY

SUBJECT ZINC FERRITE BED REGENERATION (ALL THREE GASIFIERS - HIGH TEMPERATURE GAS)

FIG. 7-1-13

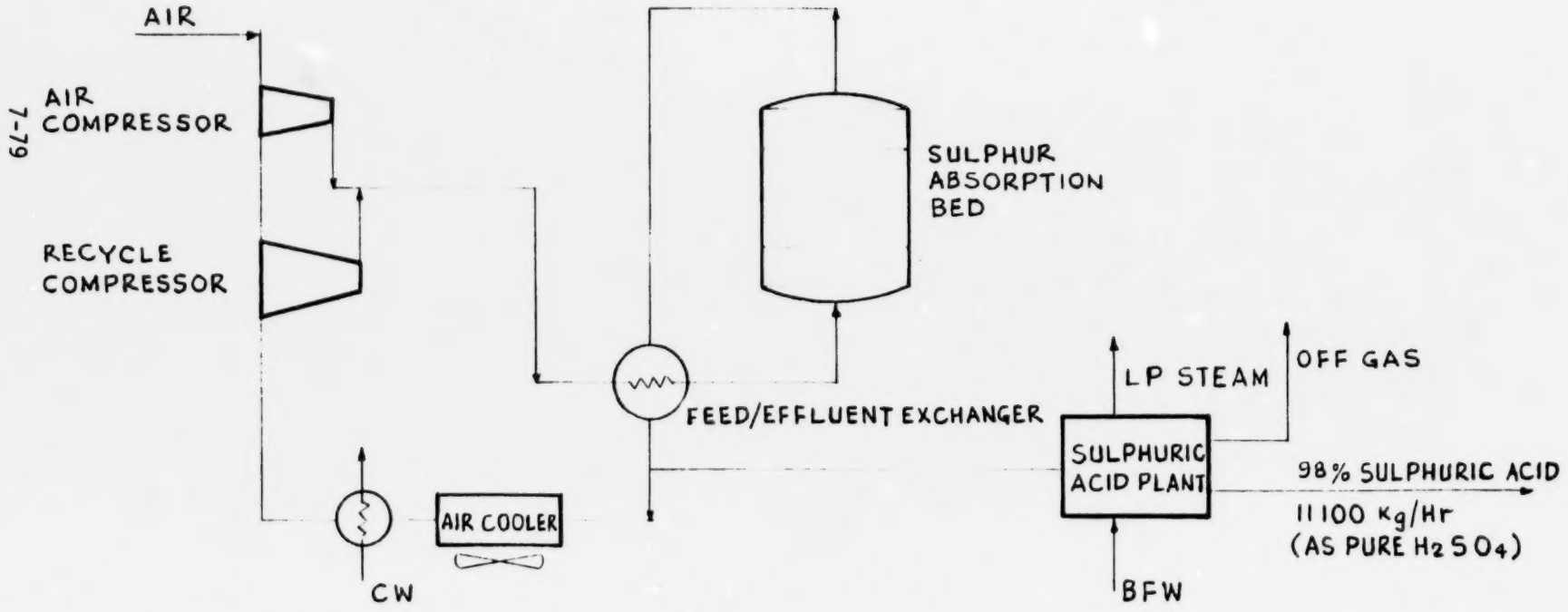

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Table 7.1-1 Case ML

PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY - LOW TEMPERATURE PURIFICATION

SUBJECT: LURGI DRY ASH OXYGEN BLOWN GASIFIER - SELEXOL PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT		Molecular Weight	(1) COAL FEED TO GASIFIER AS REQUIRED Kg/hr	Kg mol/hr	(2) OXYGEN FEED TO GASIFIER Kg/hr	Kg mol/hr	(3) ASH FROM GASIFIER Kg/hr	Kg mol/hr	(4) HYDROCARBON LIQUID & AMMONIA RECOVERY Kg/hr	Kg mol/hr	(5) RAW GAS FROM COOLING TO SULPHUR REMOVAL Kg/hr	Kg mol/hr
Methane	CH_4	16.04									12,442	775.7
Ethylene	C_2H_4	28.05									646	23.0
Ethane	C_2H_6	30.07									1,060	35.3
Hydrogen	H_2	2.02	5103	2526.2							7,642	3783.2
Carbon Monoxide	CO	28.01									38,033	1357.8
Carbon Dioxide	CO_2	44.01									121,876	2769.3
Hydrogen Sulphide	H_2S	34.08									3,634	106.6
Carbonyl Sulphide	COS	60.08									280	4.7
Nitrogen	N_2	28.02	1198	42.8	828	29.6					932	33.3
Ammonia	NH_3	17.03										
Water	H_2O	18.02	4200	233.1					1238	72.7		
Sulphur Dioxide	SO_2	64.06									580	32.2
Ash		-	9600	-			9,600					
Naptha, Tars, Oils		-										
Phenols, Others		-							5879			
Carbon, C		12.01	66598	5545.2					1095			
Oxygen,	O_2	32.00	9603	300.1	46668	1458.4	400	33.3				
Sulphur	S	32.06	3698	115.3								
TOTAL			100,000	8762.7	47,496	1488.0	10,000	33.3	8212	72.7	187,125	8921.1

EBASTEC Lavalin Inc.

Table 7.1-1 (Continued) Case ML

PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY - LOW TEMPERATURE PURIFICATION

SUBJECT: LURGI DRY ASH OXYGEN BLOWN GASIFIER - SELEXOL PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT		Molecular Weight	(6) PURIFIED FUEL GAS Kg/hr	(6) PURIFIED FUEL GAS Kg mol/hr	(7) SULPHUR RICH GAS TO SULPHUR RECOVERY PLANT Kg/hr	(7) SULPHUR RICH GAS TO SULPHUR RECOVERY PLANT Kg mol/hr	(8) TAIL GAS TREATMENT VENT GAS Kg/hr	(8) TAIL GAS TREATMENT VENT GAS Kg mol/hr	(9) ELEMENTAL SULPHUR FROM SULPHUR RECOVERY Kg/hr	(9) ELEMENTAL SULPHUR FROM SULPHUR RECOVERY Kg mol/hr
Methane	CH_4	16.04	12,374	771.5	68	4.2	32	2.0		
Ethylene	C_2H_4	28.05	590	21.0	56	2.0	28	1.0		
Ethane	C_2H_6	30.07	1,028	34.2	32	1.1	17	0.6		
Hydrogen	H_2	2.02	7,640	3782.2	2	1.0	5	2.5		
Carbon Monoxide	CO	28.01	37,984	1356.1	49	1.7	5	0.2		
Carbon Dioxide	CO_2	44.01	97,996	2226.7	23880	342.6	25069	569.6		
Hydrogen Sulphide	H_2S	34.08	140	4.1	3494	102.5				
Carbonyl Sulphide	COS	60.08	200	3.4	80	1.3	2			
Nitrogen	N_2	28.02	932	33.3			7112	253.8		
Ammonia	NH_3	17.03								
Water	H_2O	18.02	181	10.0	399	22.2	1794	99.6		
Sulphur Dioxide	SO_2	64.06								
Ash		-								
Naptha, Tars, Oils		-								
Phenols, Others		-								
Carbon, C		12.01								
Oxygen,	O_2	32.00								
Sulphur	S	32.06							3329	103.8
TOTAL			159,065	8242.5	28060	678.6	34064	929.3	3329	103.8

EBASTEC LAVALIN INC.
TABLE 7.1-3 CASE MH
PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY
SUBJECT: LURGI DRY ASH OXYGEN BLOWN GASIFIER - HIGH TEMPERATURE PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT	MOLECULAR WEIGHT	(1) COAL FEED TO GASIFIER (AS RECEIVED)		(2) OXYGEN FEED TO GASIFIER		(3) ASH FROM GASIFIER		(4) RAW GAS TO SULPHUR ABSORPTION BED		(5) PURIFIED GAS FROM SULPHUR ABSORPTION BED	
		Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr
Methane	16.04							12 442	755.7	12 442	775.7
Hydrogen	2.02	5 103	2 526.2					7 642	3 783.2	8 663	4 288.6
Carbon Monoxide	28.01							38 032	1 357.8	23 164	827.0
Carbon Dioxide	44.01							121 877	2 769.3	145 444	3 304.8
Hydrogen Sulphide	34.08							3 633	106.6		
Carbonyl Sulphide	60.08							282	4.7		
Nitrogen	28.02	1 198	42.8	566	20.2			670	23.9	670	23.9
Argon	39.95			376	9.4			376	9.4	376	9.4
Ammonia	17.03							1 238	72.7	1 238	72.7
Water	18.02	4 200	233.1					164 815	9 146.2	157 628	8 747.4
Ash	-	9 600	-			9 600					
Carbon	12.01	66 598	5 545.2			400					
Oxygen	32.00	9 603	300.1	46 668	1 458.4						
Sulphur	32.06	3 698	115.3								
Naphtha, Tars, Oils	-							5 933		5 879	
Phenols, Others	-							1 105		1 095	
Ethylene	28.05								23.0		23.0
Ethane	30.07								35.3		35.3
TOTAL		100 000	8 762.7	47 610	1 488.0	10 000		358 045	18 107.8	356 599	18 107.8

EBASTEC Lavalin Inc.

Table 7.1-5 Case FL

PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY - LOW TEMPERATURE PURIFICATION

SUBJECT: WESTINGHOUSE OXYGEN BLOWN GASIFIER - SELEXOL PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT		Molecular Weight	(1) COAL FEED TO DRYING & PREPARATION AS REQUIRED Kg/hr	Kg mol/hr	(2) OXYGEN FEED TO GASIFIER Kg/hr	Kg mol/hr	(3) ASH FROM GASIFIER Kg mol/hr	(4) RAW GAS FROM COOLING TO SULPHUR REMOVAL Kg/hr	Kg mol/hr
Methane	CH_4	16.04						10,540	657.1
Hydrogen	H_2	2.02	5,103	2526.2				5,309	2628.2
Carbon Monoxide	CO	28.01						94,976	3390.8
Carbon Dioxide	CO_2	44.01						56,051	1273.6
Hydrogen Sulphide	H_2S	34.08						3,415	100.2
Carbonyl Sulphur	COS	60.08						318	5.3
Nitrogen	N_2	28.02	1,198	42.8	540	19.3		866	30.9
Argon	Ar	39.95			385	9.6		385	9.6
Ammonia	NH_3	17.03							
Water	H_2O	18.02	4,200	233.1				292	16.2
Sulphur Dioxide	SO_2	64.06							
Ash		-	9,600				9,408		
Carbon	C	12.01	66,598	5545.2			1,262		
Oxygen	O_2	32.00	9,603	300.1	60,719	1897.5			
Sulphur	S	32.06	3,698	115.3					
TOTAL			100,000	8762.7	61,644	1926.4	10,670	172,152	8111.9

EBASTEC Lavalin Inc.

Table 7.1-5 (Continued) Case FL

PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY - LOW TEMPERATURE PURIFICATION

SUBJECT: WESTINGHOUSE OXYGEN BLOWN GASIFIER - SELEXOL PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT		Molecular Weight	(5) PURIFIED FUEL GAS Kg/hr	Kg mol/hr	(6) SULPHUR RICH GAS TO SULPHUR RECOVERY PLANT Kg/hr	Kg mol/hr	(7) TAIL GAS TREATMENT VENT GAS Kg/hr	Kg mol/hr	(8) ELEMENTAL SULPHUR FROM SULPHUR RECOVERY Kg/hr	Kg mol/hr
Methane	CH_4	16.04	10,486	653.7	54	3.4	26	1.6		
Hydrogen	H_2	2.02	5,308	2627.7	1	0.5	2	1.2		
Carbon Monoxide	CO	28.01	94,872	3387.1	104	3.7	30	1.1		
Carbon Dioxide	CO_2	44.01	46,220	1050.2	9831	223.4	13505	306.4		
Hydrogen Sulphide	H_2S	34.08	271	8.0	3144	92.2				
Carbonyl Sulphur	COS	60.08	205	3.4	113	1.9	2			
Nitrogen	N_2	28.02	866	30.9			6605	235.7		
Argon	Ar	39.95	385	9.6			1			
Ammonia	NH_3	17.03								
Water	H_2O	18.02	111	6.2	181	10.0	1182	65.6		
Sulphur Dioxide	SO_2	64.06								
Ash		-								
Carbon	C	12.01								
Oxygen	O_2	32.00								
Sulphur	S	32.06							3332	103.9
TOTAL			158,724	7776.8	13428	335.1	21,353	612.1	3332	103.9

EBASTEC LAVALIN INC.
TABLE 7.1-7 CASE FH
PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY
SUBJECT: WESTINGHOUSE OXYGEN BLOWN GASIFIER - HIGH TEMPERATURE PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT	MOLECULAR WEIGHT	(1) COAL FEED TO DRYING AND PREPARATION (AS RECEIVED)		(2) OXYGEN FEED TO GASIFIER		(3) ASH FROM GASIFIER		(4) RAW GAS TO SULPHUR ABSORPTION BED		(5) PURIFIED GAS FROM SULPHUR ABSORPTION BED	
		Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr
Methane	16.04							10 540	657.1	10 540	657.1
Hydrogen	2.02	5 103	2 526.2					5 309	2 628.2	6 255	3 096.3
Carbon Monoxide	28.01							95 349	3 404.1	81 509	2 910.0
Carbon Dioxide	44.01							57 944	1 316.6	79 940	1 816.4
Hydrogen Sulphide	34.08							3 701	108.6		
Carbonyl Sulphide	60.08							342	5.7		
Nitrogen	28.02	1 198	42.8	540	19.3			866	30.9	866	30.9
Argon	39.95			385	9.6			384	9.6	384	9.6
Ammonia	17.03							165	9.7	165	9.7
Water	18.02	4 200	233.1					33 984	1 885.9	27 506	1 526.4
Ash	-	9 600	-			9 600					
Carbon	12.01	66 598	5 545.2			1 287					
Oxygen	32.00	9 603	300.1	60 719	1 897.5						
Sulphur	32.06	3 698	115.3								
TOTAL		100 000	8 762.7	61 644	1 926.4	10 887		208 584	10 056.4	207 165	10 056.4

EBASTEC Lavalin Inc.

Table 7.1-9 Case EL

PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY - LOW TEMPERATURE PURIFICATION

SUBJECT: TEXACO OXYGEN BLOWN GASIFIER - SELEXOL PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT		Molecular Weight	(1) COAL FEED TO SLURRY PREPARATION AS REQUIRED Kg/hr	Kg mole/hr	(2) OXYGEN FEED TO GASIFIER Kg/hr	Kg mole/hr	(3) SLAG FROM GASIFIER Kg/hr	(4) RAW GAS FROM COOLING TO SULPHUR REMOVAL Kg/hr	Kg mole/hr
Methane	CH_4	16.04						139	8.7
Hydrogen	H_2	2.02	5103	2526.2				6334	3135.6
Carbon Monoxide	CO	28.01						128520	4588.4
Carbon Dioxide	CO_2	44.01						41428	941.3
Hydrogen Sulphide	H_2S	34.08						3709	108.8
Carbonyl Sulphur	COS	60.08						391	6.5
Nitrogen	N_2	28.02	1198	42.8	1109	39.6		2239	79.9
Argon	Ar	39.95			519	13.0		519	13.0
Ammonia	NH_3	17.03							
Water	H_2O	18.02	4200	233.1				335	18.6
Sulphur Dioxide	SO_2	64.06							
Ash		-	9600				9600		
Carbon	C	12.01	66598	5545.2			NIL		
Oxygen	O_2	32.00	9603	300.1	82,162	2567.6			
Sulphur	S	32.06	3698	115.3					
TOTAL			100,000	8767.7	83,790	2620.2	9600	83,614	8900.8

EBASTEC Lavalin Inc.

Table 7.1-9 (Continued) Case EL

PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY - LOW TEMPERATURE PURIFICATION (Continued)

SUBJECT: TEXACO OXYGEN BLOWN GASIFIER - SELEXOL PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT		Molecular Weight	(5) PURIFIED FUEL GAS Kg/hr	Kg mole/hr	(6) SULPHUR RICH GAS TO SULPHUR RECOVERY PLANT Kg/hr	Kg mole/hr	(7) TAIL GAS TREATMENT VENT GAS Kg/hr	Kg mole/hr	(8) ELEMENTAL SULPHUR FROM SULPHUR RECOVERY Kg/hr	Kg mole/hr
Methane	CH_4	16.04	138	8.6	1	0.1	1			
Hydrogen	H_2	2.02	6332	3134.6	2	1.0	3	1.7		
Carbon Monoxide	CO	28.01	128397	4584.0	123	4.4	12	0.4		
Carbon Dioxide	CO_2	44.01	34354	780.6	7074	160.7	7905	179.6		
Hydrogen Sulphide	H_2S	34.08	248	7.3	3461	101.6				
CarbonylSulphur	COS	60.08	250	4.2	141	2.3	2			
Nitrogen	N_2	28.02	2237	79.8	2	0.1	6160	219.8		
Argon	Ar	39.95	519	13.0			1			
Ammonia	NH_3	17.03								
Water	H_20	18.02	128	7.10	207	11.5	872	48.4		
Sulphur Dioxide	SO_2	64.06								
Ash										
Carbon	C	12.01								
Oxygen	O_2	32.00								
Sulphur	S	32.06							3329	103.8
TOTAL			172,603	8619.2	11011	281.7	14,956	449.9	3329	103.8

EBASTEC LAVALIN INC.
TABLE 7.1-11 CASE EH
PROJECT: HIGH TEMPERATURE GAS CLEAN UP STUDY
SUBJECT: TEXACO OXYGEN BLOWN GASIFIER - HIGH TEMPERATURE PURIFICATION - COMBINED CYCLE PLANT MATERIAL BALANCE

GAS COMPONENT	MOLECULAR WEIGHT	(1) COAL FEED TO GASIFIER (AS RECEIVED)		(2) OXYGEN FEED TO GASIFIER		(3) SLAG FROM GASIFIER		(4) RAW GAS TO SULPHUR ABSORPTION BED		(5) PURIFIED GAS FROM SULPHUR ABSORPTION BED	
		Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr	Kg/Hr	Kgmol/Hr
Methane	16.04							140	8.7	140	8.7
Hydrogen	2.02	5 103	2 526.2					6 284	3 110.9	7 669	3 796.3
Carbon Monoxide	28.01							128 521	4 588.4	108 589	3 876.8
Carbon Dioxide	44.01							41 427	941.3	73 030	1 659.4
Hydrogen Sulphide	34.08							3 708	108.8		
Carbonyl Sulphide	60.08							391	6.5		
Nitrogen	28.02	1 198	42.8	1 109	39.6			2 006	71.6	2 006	71.6
Argon	39.95			519	13.0			519	13.0	519	13.0
Ammonia	17.03							364	21.4	364	21.4
Water	18.02	4 200	233.1					34 816	1 932.1	24 426	1 355.5
Ash	-	9 600	-			9 600					
Carbon	12.01	66 598	5 545.2								
Oxygen	32.00	9 603	300.1	82 162	2 567.6						
Sulphur	32.06	3 698	115.3								
TOTAL		100 000	8 762.7	83 790	2 620.2	9 600		218 176	10 802.7	216 743	10 802.7

8. PROCESS AND ECONOMIC COMPARISONS

8.1 Process Comparisons

8.1.1 Thermal Efficiency

The calculated overall thermal efficiency of the coal gasification combined cycle (CGCC) plants (See Table 8.1-1) using high temperature purification shows improvement over those using low temperature purification. This is true for all three of the different modes of gasification used in the CGCC plants. The improvement in efficiency, defined as follows, is also reflected in the heat rates determined in this Study:

$$\text{Thermal Efficiency, \%} = \frac{(\text{Delivered Kw•h})\ (3600)\ (100)}{(\text{Coal, Kg/h})\quad(\text{Coal HHV kJ/kg})}$$

The degree by which the efficiency is improved by high temperature purification is not, however, the same for the three different gasification technologies. The greatest degree of improvement is shown for those incorporating moving bed gasification operating at low temperature. An improvement (decrease) of almost 25 percent is shown in the heat rate for such gasifiers (Lurgi, dry bottom) with high temperature purification. The higher temperature gasification technologies reviewed (moderate temperature fluidized bed and high temperature entrained flow) both showed about the same degree improvement or less than 10 percent.

The purpose of this Study is to investigate the overall effect resulting from the use of high temperature processes currently being considered for cleanup of raw gases generated in coal gasifiers. Thermal efficiencies represent a major area of effect. Three types of coal gasification were investigated in order to arrive at a broad range of the effect of high temperature purification. It is not the interest of this Study to compare different gasifi-

cation systems, and to do so may result in erroneous conclusions because of the reasons discussed in the following paragraphs.

For each of the gasification technologies included in this Study, the operating conditions and the expected performance used were essentially those published by EPRI (for Lurgi and Texaco gasifiers) and Westinghouse. The same conditions and performance were used with purification at high temperature as well as at low. Similarly, other features (such as steam conditions) were when practical, held the same in both the high and low temperature purification cases for each gasification technology.

It is felt that this approach and use of mathematical models to simulate turbine performance yields a reasonably reliable indication of the potential difference between specific means of hot and cold purification for a given gasification technology, although the absolute calculated performance may not be precise.

There is no pretension that the calculated performance for any of the cases is optimal, neither for the gasification technology, for the purification technology, nor for the particular gasification - purification combination. Therefore, in relation to the different gasification technologies included this Study, attention should be focused mainly on comparing hot and cold purification for a given gasifier.

The comparative thermal efficiencies in Table 8.1-1 indicate that the greatest thermal benefit would be derived by applying high temperature purification to low temperature gasification as practiced in the moving bed Lurgi dry bottom gasifier. The improvements shown for higher temperature gasification technologies studies would appear to be marginal, particularly in consideration of the very preliminary state of development of high temperature purification technology.

For all three technologies the high temperature desulfurization beds operate at approximately 600-700°C. In the case of Texaco, and to a lesser extent Westinghouse, a significant efficiency penalty is incurred in cooling the gas to this temperature. The absorption capacity of the desulfurizing materials considered fall off rapidly above the operating temperature selected, and operating temperatures become uneconomical. Additional research and development, with the aim of increasing this temperature, would appear worthy of serious consideration. Further development of the high temperature particulate removal systems and gas turbine combustion equipment may also be necessary at higher temperatures.

There are two major reasons why high temperature purification results in a large advantage when wed with the low temperature moving bed gasification technology of Lurgi, while it results in only a marginal benefit for the otner technologies.

TABLE 8.1-1

COMPARATIVE THERMAL EFFICIENCIES

FOR

LOW VERSUS HIGH TEMPERATURE PURIFICATION

	Purification Temperature Low	Purification Temperature High	Improvement (Percent)
Moving bed gasification at low temperature (Lurgi)			
Thermal efficiency, %	32.0	39.8	24.4
Heat rate, *KJ/kw•h	11260	9050	
Fluidized bed gasification at moderate temperatue (Westinghouse)			
Thermal efficiency, %	35.8	37.9	5.9
Heat rate, kJ/kw•h	10050	9500	
Entrained flow gasification at high temperature (Texaco)			
Thermal efficiency, %	36.4	39.2	7.7
Heat rate, kJ/kw•h	9885	9190	

* Heat rates are based on higher heating value of the coal.

1. Only the temperature gasification process produces condensible hydrocarbons. With low temperature purification (case ML) these hydrocarbons are removed from the raw gas coincidental with cooling of the gas for sulphur removal. The recovered hydrocarbons are assumed to be supplementary fired after the combustion turbine and thus they are converted to electricity at steam cycle efficiency of about 33 percent. With high temperature sulphur removal (case MH) it is expected that the hydrocarbons formed, including the tars, pass unchanged in vapour form through the particulate removal filters and the beds of solid sorbent. Consequently, in case MH, they are converted to electricity at a combined cycle efficiency of over 45 percent.

2. The low temperature Lurgi dry bottom gasifier utilizes a large quantity of steam principally to moderate the temperature and avoid clinkering of the ash. As a result, the raw gas leaving the gasifier proper is about one-third steam. With low temperture sulphur removal most of the high level heat (above 190°C) is removed, but not recovered, in the quench scrubber which is used as an intergral part of the Lurgi process to remove particulate and to condense and washout the tars. With high temperature purification there is conceptually no need for such quenching and the sensible heat of the steam and the high level heat of the gases are converted to electricity in the combined cycle plant.

   The raw gases leaving the higher temperature gasifiers (Westinghouse and Texaco) also contain substantial quantities of steam. Witnout the presence of tars and other high boiling hydrocarbons, these technologies when used with low temperature purification recover the high level sensible heat in boilers before quenching to remove particulate matter. Consequently, with these gasifiers the change in heat usefully recoverable that comes with adoption of high temperature purification is not as drastic as with the Lurgi gasifier.

8.1.2 Technical Risk

It is the purpose of this Study to determine the projected change in the cost of producing electricty when the purification of the coal-derived gases is accomplished at high temperature. Others will decide if the predicted change is sufficient either to motivate continued interest in the development and future commercialization of hot purification, or to dismiss further consideration.

Such a decision involves risk which should be recognized. The predicted changes in the cost of electricity reflect estimated operating requirements as well as estimated capital needs. One risk involved is whether the projected differences between hot and cold purification for a given gasifier are accurate as well as consistent. They are felt to be reasonably consistent because tangible steps could be and were taken to do so. The inaccuracy, if such exists, is primarily with the cost predictions using hot purification and is relatively minor with those using cold purification.

The high temperature sulphur removal facilities used with all the gasification technologies is based on published reports of the Morgantown Energy Research Centre (MERC). Unfortunately, the work reported on was of a preliminary screening nature at bench scale and with simulated gases. Consequently, the conceputal process design used in this Study must be considered only order-of-magnitude. The operating and capital costs based thereon must in turn be appraised as being soft.

These comments are even more applicable to the facilities for regeneration of the sulphided sorbent since with such a highly exothermic reaction extensive scaleup is difficult when control of temperatures is critical.

The high temperature alternates are envisioned to require more than just the use of a sorbent to remove the sulphur and the means to regenerate the sorbent

and recover the sulphur. It is envisioned to also require particulate removal before the sorbent so as to avoid plugging, and after the sorbent to protect the combustion turbine from sorbent degradation, and to protect the turbine from harmful metallic contaminants. While considerable exploratory work has been done in these areas, most of it to date has been associated with the fluidized combustion of coal. Whether such data is equally valid in a reducing atmosphere as in an oxidizing atmosphere is questionable. This factor as well as the limited nature of the information available, made even a conceptual design of such facilities impractical.

If the projected change in the cost of electricity using high temperature purification is sufficient there is considerable development and investigative work to be done in order to reduce the technical risks to the level existing with low temperature purification. This work would be needed to provide answers to many questions of which the following are but a few:

a) What is the nature of the particulate matter which exists in the raw gas leaving the gasifier proper? What is its quantity and size distribution?

b) How are the characteristics of the particulate matter leaving the gasifier influenced by the gasifier operating conditions?

c) What are the requirements of the sulphur sorbent with respect to particulate matter in the gas? Is the entering particulate matter retained by the sorbent and to what extent? Is that which is retained, removable? How?

d) If means are required to remove particulate matter before sulphur removal what means are effective? Are they damaged by impurities and contaminants in the raw gas?

e) Is the capacity of the sorbent and its effectiveness dependent upon the sulphur compounds in the entering gas? How?

f) Is the sorbent damaged by impurities and contaminants in the raw gas?

g) What is the life or the number of cycles the sorbent is capable of sustaining without intolerable loss of effectiveness?

h) What metallic and alkali contaminants exist in the raw gas from the gasifier? Is their nature and extent predictable on the basis of the operating conditions and the coat?

i) Are there effective means for removing such contaminants to levels compatible with desired turbine life?

8.1.3 Environmental Impact

This Study did not attempt to conform to particular specific legal requirements with respect to dust emissions, solids disposal, liquid effluents, or gaseous emissions. To do so, would require abandoning the general nature of the Study, and making it site specific. The approach used avoids prejudicing the evaluation of the purification means by use of a requirement not reasonably achievable by one or the other purification means, and which would not be necessary in many possible locations.

The following discussion of the environmental impact essentially addresses whether hot purification will tend to make meeting specific environmental requirements easier or more difficult than will low temperature purification.

8.1.3.1 Dust Emissions

This Study did not attempt to conform to particular specific legal requirements with respect to dust emissions, solids disposal, liquid effluents, or gaseous nature of the Study, and making it site specific. The approach used avoids prejudicing the evaluation of the purification means by use of a requirement not reasonably achievable by one or the othe purification means, and which would not be necessary in many possible locations.

The following discussion of the environmental impact essentially addresses whether hot purification will tend to make meeting specific environmental requirements easier or more difficult than will low temperature purification.

The level of fugitive coal dust emissions for a CGCC plant using a given gasification technology would basically be the same whether cold or hot purification was used. This parity is altered somewhat by the fact that the purification technology having the highest gasifier heat rate would have to process more coal, and would, therefore, emit more coal dust over a given time period.

With high temperature purification zinc ferrite dust potentially may be periodically emitted to the environment when loading and unloading the sulphur absorption vessels. The level and quantity would be largely dependent upon the friability of the sorbent material and upon its useful life.

8.1.3.2 Liquid Effluents

The use of high temperature purification would essentially eliminate liquid process effluents from the plant, and thus negate the need for treatment to make them environmentally acceptable. The ultimate source of the bulk of the

liquid effluents from a CGCC plant using low temperature purification are the water vapours and condensible hydrocarbons in the raw gas as well as the water introduced for quenching and scrubbing of the gas before sulphur removal. With high temperature purification the water vapors from the hydrocarbons in the raw gas pass into the combustion turbine and are emitted from the plant essentially as CO_2 and water vapor in the stack gases.

8.1.3.3 Gaseous Emissions, Sulphur Dioxide

The recovery of at least 90 percent of the sulphur contained in the coal gasified is the basis used for this Study. Assuming no signficant sulphur with the ash or slag this establishes the total discharge to the atmosphere at not over 0.26 mg of SO_2 per kilojoule (HHV) of coal gasified. The bulk of this emission is in the stack gases discharged from the combustion turbine heat recovery facilities.

Both the high and low temperature sulphur removal means used for this Study are capable of such performance without difficulty. Imposition of increasingly stricter emission requirements would tend in the case of low temperature purification (Selexol) to require taking additional measures to achieve them. These would include use of refrigeration to cool the lean solvent and the catalytic hydrolysis of carbonyl sulphide to permit its easier removal as hydrogen sulphide. Case ML with the low temperature Lurgi gasifier would have the most difficult time to reach low emission levels because about 3 to 5 percent of the sulphur appears in the liquid hydrocarbon byproducts and is therefore not subject to removal by Selexol. It is anticipated that in a CGCC plant these hydrocarbons would be fired in the combustion turbine heat recovery train without sulphur removal before or after their burning.

The design of the high temperature sulphur removal facilities of this Study conceptually would remove essentially all of the sulphur from the raw gas assuming it was all hydrocarbon sulphide. The work at METC with zinc ferrite

did not deal with other sulphur compounds known to exist in some gasifier effluent gases. In the chemical industry, it has long been recognized that zinc oxide is not as effective for removal of such compounds as it is for hydrogen sulphide. Since only about 90 to 95 percent of the sulphur is normally present as hydrogen sulphide there may be limits on the practical level to which sulphur dioxide emissions can be reduced using only zinc ferrite. However, high temperature purification using zinc ferrite would appear better able to meet very stringent emission requirements than would the current low temperature processes.

8.1.3.4 Gaseous Emissions, Nitrogen Oxides

The meeting of nitrogen oxide emission standards will tend to be more difficult with high temperature purification than with purification at low temperature. Two factors contributed to this. First, as presently envisioned, the nitrogen compounds such as ammonia in the raw gas would pass through to the combusion turbine unchanged. The cooling of the raw gas which is necessary for low temperature sulphur removal makes almost complete removal of these compounds possible.

Second, the high temperature of the fired gas will tend to result in higher reaction zone temperature in the combustor unless more of the air is used for combustion with less for cooling and dilution, or unless steam or water is used to moderate reaction zone temperature.

To reduce the level of NOx for low bound-nitrogen fuels the injection of steam has been favoured for combined cycle plants. However, the evidence to date indicates that steam injection is not effective with high bound-nitrogen fuels, particularly coal-derived fuels, therefore advanced dry, low NO_x combustion techniques must be developed. Several advanced low NOx conceptual design approaches have been identified by Westinghouse under the NASA low

No_x Heavy Fuel Combustion Concept Program NASA Den 3-146 and the most applicable of these is the Rich-Lean Combustor (Reference 1).

Rich-Lean Combustion is a two stage process in which the initial rich stage converts fuel bound nitrogen to N_2 and essentially no thermal NOx is produced. The second lean stage completes the combustion process with negligible Nox generation. This approach offers low emission compatibility with high bound-nitrogen fuels.

8.1.3.5 Gaseous Emissions, Particulate Matter

Without doubt the meeting of particulate emissions limitations is more difficult with high temperature purification. The quenching and scrubbing of the gases practiced with all gasification technologies before low temperature sulphur removal is inherently capable of removing the particulate matter. With high temperature sulphur removal, however, the sorbent used is in itself a potential source of dust in the fuel and hence in the stack gas.

The following attempts, to quantify, based on the limited information available, the sulphur compounds, NOx and particulates in the fuel gas for the six cases.

8.1.3.6 Lurgi Moving Bed Gasifier - Case ML

The raw gas analysis, Section 4.2.1 indicates that there is 1.64 percent by weight of naptha, tars and oils, and 0.30 percent by weight of phenols and others in the gas. The particulate matter is approximately 0.2 kg/m^3. The gas contains sulphur in the form of H_2S, 0.59 Mole percent and COS, 0.03 Mole percent. The raw gas is passed through a scrubber where it is washed by circulating gas liquor and the particulate and tars removed. It is expected that the scrubbing process removes particulate in fuel gas to less than 30 ppm by weight which is an acceptable level for General Electric Gas Turbines. The

Lurgi process removes the phenols and ammonia in the gas liquor leaving an effluent with amounts below 20 ppm phenols and below 60 ppm free ammonia. This is further processed to meet regulations for the disposal of waste water.

The Selexol sulphur removal plant removes 93.3 percent of the fuel gas sulphur and the Claus and Beavon units recover 99.97 percent of this in incremental form. This produces a purified fuel gas Table 7.1-1 containing 4.1 kg mol/h H_2S and 3.4 kg mol/h COS based on 100 metric tonnes per hour of coal and an output of 252.69 Megawatts. This is equivalent at the plant power output of 215 Megawatts to 0.05 percent H_2S plus 0.04 percent COS or a total of 900 PPM. These levels are above those quoted in Section 5 by Westinghouse for high reliability gas turbines but information is not available on the effect of this on the first stage blade life. The low levels of the sulphur gases and nitrogen in the fuel gas will be considerably diluted during combustion in the gas turbine, producing low levels of SO_2 and NO_x that are in compliance with Federal emission control guidelines.

8.1.3.7 Lurgi Moving Bed Gasifier - Case MH

The raw gas analysis of Section 4.2.1, given in Case ML applies. Sulphur removal from the raw medium B.T.U. gas is achieved through high temperature absorption using zinc ferrite. Most of the dust in the gas is removed by the cyclone system. The remaining particulate matter is removed by ceramic fiore filters. These would be designed to limit the solid materials in the fuel gas to well below the 30 PPM by weight, the maximum acceptable level for General Electrical Gas Turbines. This will probably be above the 0.06 grains/scf (1 ppm) recommended by Westinghouse. The gas enters the zine ferrite sulphur absorption bed at about 580°C. The high temperature sulphur removal zinc ferrite bed removes practically all the hydrogen sulphide and carbonyl sulphide.

This produces a purified fuel gas Table 7.1-3 containing very little H_2S and COS. The high level of ammonia and nitrogen in the purified gas which normally convert to NO_x in the gas turbine will not be effectively reduced by the high steam content. To reduce the high levels of NOx produced from the high bound-nitrogen fuel it is necessary to use advanced type low NOx combustors that are being developed. It is expected that the NO_x formed in combustion will be effectively reduced by advanced type combustors to provide compliance with Federal emission control guidelines.

8.1.3.8 Westinghouse Fluidized Bed Gasifier - Case FL

The gasifier characteristics Table 4-1, indicate that there are only trace amounts of naphtha, tars and oils in the gas. The particulate matter is approximately 0.4 kg/m^3. The gas contains sulphur to the extent of 1.14 Mole percent H_2S plus COS.

The fines laden raw gas from the gasifier flows through two cyclones, which separate between 80 to 90 percent of the elutriated char. Sensible heat is recovered from the product gas in high, medium, and low pressure waste heat boilers which increases overall process efficiency. The gas exiting the low pressure boiler is further cooled and scrubbed of any remaining particulate matter entrained in the gas by a direct water quench venturi scrubbing system. It is expected that the scrubbing process reduces the particulate level to less than the 30 ppm by weight which is the acceptable level for General Electric Gas Turbines.

The water-fines slurry from the scrubber bottoms is processed through a clarifier pit.

The Selexol sulphur removal plant removes 90.1 percent of the fuel gas sulphur and the Claus and Beavon units recover 99.97 percent of this in elemental form. This produces a purified fuel gas Table 7.1-5 containing 8.0 kg mol/h

H_2S and 3.4 kg mol/h COS based on 100 metric per hour of coal and an output of 252.69 Megatwatts. This is equivalent at the plant power output of 215 Megawatts to 0.10 percent H_2S plus 0.044 percent COS or a total of 1466 ppm. These levels are above those quoted by Westinghouse in Section 5 but information is not available on the effect of this on the first stage blade life.

The emissions as stated for the Lurgi ML case will be in compliance with Federal guidelines.

8.1.3.9 Westinghouse Fluidized Bed Gasifier - Case FH

The raw gas analysis of Section 4.2.2, given in case FL applies. Sulphur removal from the raw medium B.T.U. gas is achieved through high temperature absorption using zinc ferrite. Most of the dust in the gas is removed by the cyclone system. The remaining particulate matter is removed by ceramic fibre filters. These would be designed to limit the solid materials in the fuel gas well below the 30 ppm by weight, the maximum acceptable level for General Electric Gas Turbines.

This produces a purified fuel gas Table 7.1-7 containing only trace amounts of H_2S and COS. The gas leaves the zinc ferrite sulphur absorption level to less than 30 ppm by weight which is the acceptable level for General Electric Gas Turbines.

8.1.3.10 Texaco Entrained Bed Gasifier - Case EL

The raw gas analysis, Section 4.2.3 indicates that there are no naphtha, tars tars and oils due to the high temperature of the gasifier. The particulate matter is approximately 0.7 kg/m^3. The gas contains sulphur in the form 1.01 Mole percent H_2S and 0.06 Mole percent COS.

The fines laden hot fuel gas issuing from the gasifier is cooled by the radiant and convection coolers in the quench tower, generating high pressure (10,470 kPa) saturated steam in the process. The temperature of the gas is well below the ash softening point thereby solidifying the entrained ash particles. The gas then enters the carbon scrubber where most of the fine particulate material is removed. Counter current contact with the water captures entrained particles to produce a particulate free saturated gas at a temperature of 178°C. The synthesis gas leaves the quench tower and is further cooled in a gas cooling train to 38°C.

It is expected that the scrubbing process reduces the particulate level to less than the 30 ppm by weight which is the acceptable level for General Electric Gas Turbines.

The Selexol sulphur removal plant removes approximately 90 percent of the fuel gas sulphur and the Claus and Beavon units recover 99.96 percent of this in incremental form. This produces a purified fuel gas Table 7.1-9 containing 7.3 kg mol/h H_2S and 4.2 kg mol/h COS based on 100 metric tonnes per hour of coal and an output of 252.69 Megawatts. This is equivalent at the plant power output of 215 Megawatts to 0.085 percent H_2S plus 0.05 percent COS or a total of 1334 ppm. These levels are above those quoted by Westinghouse in Section 5 but information is not available on the effect of this on the first stage blade life.

The emissions as stated for the Lurgi ML case will be in compliance with Federal guidelines.

8.1.3.11 Texaco Entrained Bed Gasifier - Case EH

The raw gas analysis of Section 4.2.3 given in case EL applies. Sulphur removal from the raw medium B.T.U. gas is achieved through high temperature absorption using zinc ferrite. Most of the dust in the gas is removed by the

quench tower and scrubber system. The remaining particulate matter is removed by ceramic fibre filters. These would be designed to limit the solid materials in the fuel gas well below the 30 ppm by weight, the maximum acceptable level for General Electric Gas Turbines.

This produces a purified fuel gas Table 7.1-11 containing only a trace of H_2S and COS. The low level of ammonia in the purified gas which normally converts to NO_x in the gas turbine will not be effectively reduced by the high steam content. It is expected that the NO_x formed in combustion will be effectively reduced by advanced type combustors to provide compliance with Federal emission control guidelines.

8.2 Economic Comparisons

In the following subsection the total unit power costs at 1982 price levels are presented. This is followed by a presentation of the life cycle costs associated with the six options. The sensitivity to the life cycle costs to variation in a number of parameters is then presented.

8.2.1 Total Unit Power Costs

The capital and operating costs associated with the six process alternatives together with the underlying assumptions used in these costs are presented in Section 7.

These costs together with the average unit power costs are summarized in Table 8.2.1-1.

The costs were developed from previous cost estimates contained in Electric Power Research and U.S. Department Reports. These incorporated similar processing units and these have been factored or sealed for the smaller unit. It is concluded that because of the many assumptions, including the high uncer-

tainty in hot purification costs, the accuracy of the economics presented cannot be specified.

The estimated capital costs of Section 7.2 therefore provide a basis for the order of magnitude economics that are significant in identifying the best possible applicable for high temperature purification.

Capital cost comparisons are therefore more relevant when for each of the three gasification technologies, the hot and cold temperature purification cases are compared. Considerable less validity is attached to comparisons between gasifiers with hot and cold temperature purification.

The change in power cost by replacing the low temperature purification with high temperature purification for each gasifier reveals the following:

Lurgi Dry Bottom Gasifier
Reduction in power cost from 0.0982 to 0.0872 $/kWH

Westinghouse Fluidized Bed Gasifier
Increase in power cost from 0.0816 to 0.0900 $/kWH

Texaco Entrained Bed Gasifier
Increase in power cost from 0.0917 to 0.0977 $/kWH.

Examination of Table 8.2-1, Summary Total Unit Power Costs indicates that the Lurgi Dry Bottom Gasifier has the greatest benefit by replacing low temperature purification with high temperature purification. This can be attributed to the deletion of equipment required for cold temperature purification, and is mainly the water treatment equipment and the tar combustion unit for tar removal. The increase in thermal efficiency, Table 8.1-1 for high temperature purification increases cost savings for coal and equipment.

The Westinghouse and Texaco gasifiers both show increased power costs with high temperature purification. The improvement in thermal efficiency shown in Table 8.1-1 is also small. There is, therefore, no economic advantage in proposing high temperature purification for these gasifiers.

The unit power cost of the Lurgi gasifiers with high temperature purification, the Westinghouse gasifier with low temperature purification, and the Texaco gasifier with low temperature purification all lie within a 12 percent scatterband. This indicates that the cost of power for Lurgi with high temperature purification is probably competitive with the Westinghouse and Texaco with low temperature purification.

TABLE 8.2-1

SUMMARY TOTAL UNIT POWER COSTS

(Mid 1982 Dollars)

	LURGI		WESTINGHOUSE		TEXACO	
	Case ML	Case MH	Case FL	Case FH	Case EL	Case EH
Total Capital Required(1)	452,278	392,486	365,797	411,419	428,356	461,489
Total Net Operating Cost Per Year(1)	55,251	50,697	47,430	51,245	50,589	53,085
Total Capital Changes Per Year(1)	74,193	64,299	60,086	67,413	70,289	75,660
Total Cost of Service Per Year(1)	129,444	114,996	107,516	118,658	120,878	128,751
Annual Power Output MWH	1,318,380	1,318,380	1,318,380	1,318,380	1,318,380	1,318,380
Total Unit Power Cost ($/kWH)	0.0982	0.0872	0.0816	0.0900	0.0917	0.0977
Percent Difference Above Lowest Cost	+20.3	+6.86	0	+10.3	+12.4	+19.7

Note: (1) $ Cdn x 1000

8.2.2 Life Cycle Costs

The assumptions employed in performing the life cycle cost analyses are presented below.

Cost Estimates:	The capital and operating costs employed in the analysis are estimated at 1982 price levels in Section 7.
Relevant Capital Costs:	Total Plant Investment plus all capital charges except construction loan interest (construction loan interest is accounted for by the discounting process).
Working Capital:	The initial working capital requirements are estimated in Section 7. This item is then indexed to 1986 price levels. Additional working capital is provided each year to meet inflationary requirements.
Relevant Operating Costs:	All direct operating costs are included in the life cycle cost analysis. Depreciation is not included as it is a non cash expense. Interest on long term debt is not included as the financing changes are accounted for by the discounting procedure.
Inflation:	Both capital and operating costs are assumed to be subject to inflation at a compound rate of 6 percent per year.
Timing of Capital Expenditures:	Construction is assumed to commence in mid-1983 and be complete by mid-1986. Capital expenditures are assumed to be weighted as follows:

- 10% in 1983
- 40% in 1984
- 40% in 1985
- 10% in 1986

Business Taxes: Taxes on operating profits are assumed to be zero (i.e. the public utility case).

Plant Life: The plant is assumed to start operation in mid-1986 and operate until the end of 2010 (i.e. 25 year life).

Salvage Value: Equal to recovered working capital. It is realized at the end of the final year of operation.

Discount Rate: A discount rate of 12.5 percent is employed. This is deemed to be a reasonable estimate of the appropriate weight average cost of capital for utilities under the general scenario of 6 percent inflation.

Levelized Unit Power Cost or Average Discounted Unit Power Cost: The present value of total life cycle costs of the plant divided by the present value of the total power production during the life of the plant.

The resulting present values of the Life Cycle costs and levelized unit power costs for the six process alternatives are summarized in Table 8.2.2-1. Tables 8.2.2-1 and 8.2.2-2 provide an illustration of the analysis employed in developing these estimates.

TABLE 8.2.2-1

SUMMARY OF LIFE CYCLE COST ANALYSIS

	P.V. Life Cycle Costs (1983 $ x 1 million)	Levelized Unit Power Costs (1983 $/kWH)
Case ML	1022.2	0.1376
Case MH	920.0	0.1238
Case FL	858.2	0.1155
Case FH	942.2	0.1268
Case EL	947.7	0.1275
Case EH	1005.8	0.1354

8.2.3 Sensitivity Analysis

The variation in total life cycle cost and levelized unit power costs the changes in the following parameters is investigated.

- Plant Capacity
- Capital Cost
- Operating Cost and
- Inflation.

The results of these analyses are summarized in Tables 8.2.3-1 and 8.2.3-2. Figures 8.2.3-1 and 8.2.3-2 provide a graphical summary of the levelized power costs.

TABLE 8.2.3-1

SUMMARY OF SENSITIVITY ANALYSIS
PRESENT VALUE OF LIFE CYCLE COSTS (1983 $ x 1000)

	Case ML	Case MH	Case FL	Case FH	Case EL	Case EH
Plant Capacity						
Base Case=70%	1,022,180	919,999	858,190	942,181	947,659	1,005,821
Capacity =50%	924,356	924,776	769,821	848,777	859,661	914,984
Capacity =60%	972,522	871,659	813,330	894,765	902,990	959,711
Capacity =80%	1,070,573	967,091	901,894	988,381	991,197	1,050,767
Capital Cost						
Base Case	1,022,180	919,999	858,190	942,181	947,659	1,005,821
75% of Base	928,089	838,665	782,233	856,950	858,525	910,003
125% of Base	116.270	1,001,333	934,147	1,027.412	1,036,794	1,101,639
Operating Costs						
Base Case	1,002,180	919,999	858,190	942,181	947,654	1,005,821
75% of Base	860,722	771,080	721,026	791,495	801,733	850,012
125% of Base	1,183,637	1,068,918	995,354	1,092,867	1,093,585	1,161,629
Inflation						
Base Case	1,022,180	919,999	858,190	942,181	947,659	1,005,821
Constant $1982	644,956	575,504	536,938	592,279	600,783	639,507
1% Real Increase in Coal Cost	1,068,358	957,104	899,388	981,122	988,191	1,043,508

TABLE 8.2.3-2

SUMMARY OF SENSITIVITY ANALYSIS
LEVELIZED POWER COST (1982 $/kWH)

	Case ML	Case MH	Case FL	Case FH	Case EL	Case EH
Plant Capacity						
Base Case = 70%	0.1376	0.1238	0.1155	0.1268	0.1275	0.1354
Capacity = 50%	0.1736	0.1549	0.1446	0.1594	0.1615	0.1719
Capacity = 60%	0.1529	0.1370	0.1279	0.1407	0.1420	0.1509
Capacity = 80%	0.1264	0.1142	0.1066	0.1167	0.1170	0.1240
Capital Cost						
Base Case	0.1376	0.1238	0.1155	0.1268	0.1275	0.1354
75% of Base	0.1249	0.1129	0.1053	0.1153	0.1156	0.1225
125% of Base	0.1502	0.1348	0.1257	0.1383	0.1395	0.1483
Operating Cost						
Base Case	0.1376	0.1238	0.1155	0.1268	0.1275	0.1354
75% of Base	0.1158	0.1038	0.0970	0.1065	0.1079	0.1144
125% of Base	0.1593	0.1439	0.1340	0.1472	0.1472	0.1563
Inflation						
Base Case	0.1376	0.1238	0.1155	0.1268	0.1275	0.1354
Constant $1982	0.0868	0.0775	0.0723	0.0747	0.0809	0.0806
% Real Increase in Coal Costs	0.1438	0.1288	0.1211	0.1321	0.1330	0.1404

8.2.3 Sensitivity Analysis on the Effect of Varying the Following Parameters on Levelized Power Cost $/KWh

Table 8.2.3.2, Figure 8.2.3-1 and Figure 8.2.3-2 indicate the trends for the parameters; plant capacity, capital cost, operating cost and inflation. The lowest cost FL case is selected to depict the effect which is similar for the other five cases.

8.2.3.1 Plant Capacity

In the FL case, the base cost of 70 percent plant capacity is 0.1155 $/KWh and for 50 percent plant capacity is 0.1446 $/KWh. The power cost increases 25 percent, with the reduction in plant capacity from 70 percent to 50 percent. Similar increases are apparent for the other five cases and are probably due to the effect of the fixed costs, capital cost and operating labour on the total cost of power.

8.2.3.2 Capital Cost

In the FL case, the base cost of power is 0.1155 $/KWh and for 125 percent of base is 0.1257 $/KWh. The power cost increase is 8.8 percent for a 25 percent increase in capital cost. The relatively low power cost increase is probably attributable to the high ratio of operating costs to capital costs.

8.2.3.3 Operating Cost

In the FL case, the base cost of power is 0.1155 $/KWh and for 125 percent of base is 0.1340 $/KWh. The power cost increase is 16 percent for a 25 percent increase in operating costs. The relatively high power cost increase is probably caused by the high ratio of operating costs to capital costs.

8.2.3.4 Inflation

In the FL case, the base cost of power based on inflation rate of 6 percent compounded is 0.1155 $/KWh and for zero inflation (constant 1982 dollar) is 0.072 $/KWh. The power cost increase attributed to the 6 percent compounded inflation is 60.4 percent.

A one percent real increase in the cost of coal (over and above base case inflation assumption) causes the cost of power to rise to 0.1211 from 0.1155 $/KWh, an increase of 4.8 percent. The cost of power is therefore very sensitive to an increase in coal price.

8.2.2-2

TABLE 8.2.2-2

HIGH TEMPERATURE GAS CLEAN UP STUDY

CASE HL-BASE CASE ASSUMPTIONS
GASIFICATION: MOVING BED AT LOW TEMPERATURE
PURIFICATION: LOW TEMPERATURE SULPHUR REMOVAL

PROJECTED OPERATING COST STATEMENTS ($X1000)

	YEAR 1983	YEAR 1984	YEAR 1985	YEAR 1986	YEAR 1987	YEAR 1988	YEAR 1989	YEAR 1990	YEAR 1995	YEAR 2000	YEAR 2005	YEAR 2010
ASSUMPTIONS												
PLANT OUTPUT (MW)	0	0	0	75	151	151	151	151	151	151	151	151
FULL PLANT CAP (MW)	0	0	0	215	215	215	215	215	215	215	215	215
PERCENT CAPACITY	0	0	0	35	70	70	70	70	70	70	70	70
QUANTITY ELECT (MWH)	0	0	0	656124	1318380	1318380	1318380	1318380	1318380	1318380	1318380	1318380
GEN PRICE INF IND	1.000	1.124	1.191	1.262	1.338	1.419	1.504	1.594	2.133	2.854	3.820	5.112
OPERATING COSTS												
COAL REQUIRED	0	0	0	18657	39739	42123	44650	47329	63337	84760	113427	151791
RAW WATER REQUIRED	0	0	0	57	120	128	135	143	192	257	344	460
CATALYSTS & CHEMS	0	0	0	323	688	729	773	819	1096	1467	1963	2627
OPERATING LABOUR	0	0	0	991	2110	2237	2371	2513	3364	4501	6024	8061
ASH DISPOSAL EXP.	0	0	0	63	134	142	150	159	213	285	382	511
MAINTANCE LABOUR	0	0	0	2980	6347	6728	7132	7560	10116	13538	18117	24245
SUPERVISION	0	0	0	1191	2537	2690	2851	3022	4044	5412	7242	9692
UPP & MIC MATERIALS	0	0	0	2284	4864	5156	5466	5794	7753	10376	13885	18581
GEN & ADMIN EXPENSES	0	0	0	3973	8463	8971	9509	10079	13489	18051	24156	32326
PROPERTY TAX & INS	0	0	0	4967	10579	11213	11886	12599	16861	22564	30195	40408
CREDIT FOR SULPHUR	0	0	0	-324	-691	-732	-776	-823	-1102	-1475	-1974	-2642
CREDIT FOR AMMONIA	0	0	0	-446	-950	-1008	-1068	-1132	-1516	-2028	-2715	-3633
TOT BY-PRODUCT CRED	0	0	0	-771	-1642	-1741	-1845	-1956	-2618	-3504	-4690	-6276
GROSS OPERATING COST	0	0	0	35486	75582	80117	84924	90019	120466	161210	215735	288703
NET OPERATING COST	0	0	0	34714	73938	78375	83077	88062	117846	157705	211045	282425
DEP & AMORT	0	0	0	17658	17658	17658	17658	17658	17658	17658	17658	17658
INTEREST	0	0	0	55182	55182	55182	55182	55182	55182	55182	55182	55182
TOTAL COSTS	0	0	0	107554	146778	151215	155917	160902	190686	230545	283885	355265

8.2.2-2

TABLE 8.2.2

HIGH TEMPERATURE GAS CLEAN UP STUDY

CASE ML-BASE CASE ASSUMPTIONS
GASIFICATION: MOVING BED AT LOW TEMPERATURE
PURIFICATION: LOW TEMPERATURE SULPHUR REMOVAL

LIFE CYCLE COST ANALYSIS

	YEAR 1983	YEAR 1984	YEAR 1985	YEAR 1986	YEAR 1987	YEAR 1988	YEAR 1989	YEAR 1990	YEAR 1995	YEAR 2000	YEAR 2005	YEAR 2010
CASH FLOW												
TOTAL COSTS	0	0	0	107554	146778	151215	155917	160902	190686	230545	283885	355265
DEP & AMORT	0	0	0	-17657	-17657	-17657	-17657	-17657	-17657	-17657	-17657	-17657
INTEREST	0	0	0	-55181	-55181	-55181	-55181	-55181	-55181	-55181	-55181	-55181
CASH FLOW (EX FIN)	0	0	0	34714	73938	78375	83077	88062	117846	157705	211045	282425
POWER PRODUCED												
QUANTITY ELECT (MWH)	0	0	0	656124	1318380	1318380	1318380	1318380	1318380	1318380	1318380	1318380
DISC POWER @12.5%	0	0	0	460817	823058	731607	650318	578060	320782	178011	98784	54818
ACC DISC POWER(MWH)	0	0	0	460817	1283875	2015482	2665800	3243860	5302082	6444249	7078071	7429797
INVESTMENT ANALYSIS												
CAPITAL-LAND,P & E	41103	174277	184734	48954	0	0	0	0	0	0	0	0
WORKING CAPITAL REQD	0	0	0	13665	820	869	921	977	1307	1749	2340	3132
SALVAGE VALUE	0	0	0	0	0	0	0	0	0	0	0	-55328
CASH FLOW (EX FIN)	0	0	0	34714	73938	78375	83077	88062	117846	157705	211045	282425
NET CASH FLOW-CAP	41103	174277	184734	97334	74758	79244	83998	89038	119153	159454	213385	230228
10% DISCOUNT RATE												
DISC CASH FLOW-C1	41103	158434	152672	73128	51061	49204	47415	45691	37966	31547	26213	17561
DCF PAY BACK-C1	41103	199537	352209	425338	476398	525603	573017	618708	823415	993513	1134852	1248076
NET PRES VAL-C1	1248076	0	0	0	0	0	0	0	0	0	0	0
12.5% DISCOUNT RATE												
DISC CASH FLOW-C2	41103	154913	145962	68361	46671	43975	41434	39040	28992	21530	15989	9573
DCF PAY BACK-C2	41103	196016	341979	410339	457010	500985	542419	581459	745320	867006	957373	1022181
NET PRES VAL-C2	1022181	0	0	0	0	0	0	0	0	0	0	0
15% DISCOUNT RATE												
DISC CASH FLOW-C3	41103	151545	139685	63999	42743	39398	36315	33473	22271	14817	9859	5288
DCF PAY BACK-C3	41103	192648	332334	396332	439075	478474	514788	548261	680198	767980	826384	863972
NET PRES VAL-C3	863972	0	0	0	0	0	0	0	0	0	0	0
RESULT												
LEVELIZED COST $/KWH	0.137e											

FIGURE No 8.2.3-1
GRAPHICAL SUMMARY OF SENSITIVITY ANALYSIS
LOW TEMPERATURE PURIFICATION OPTIONS

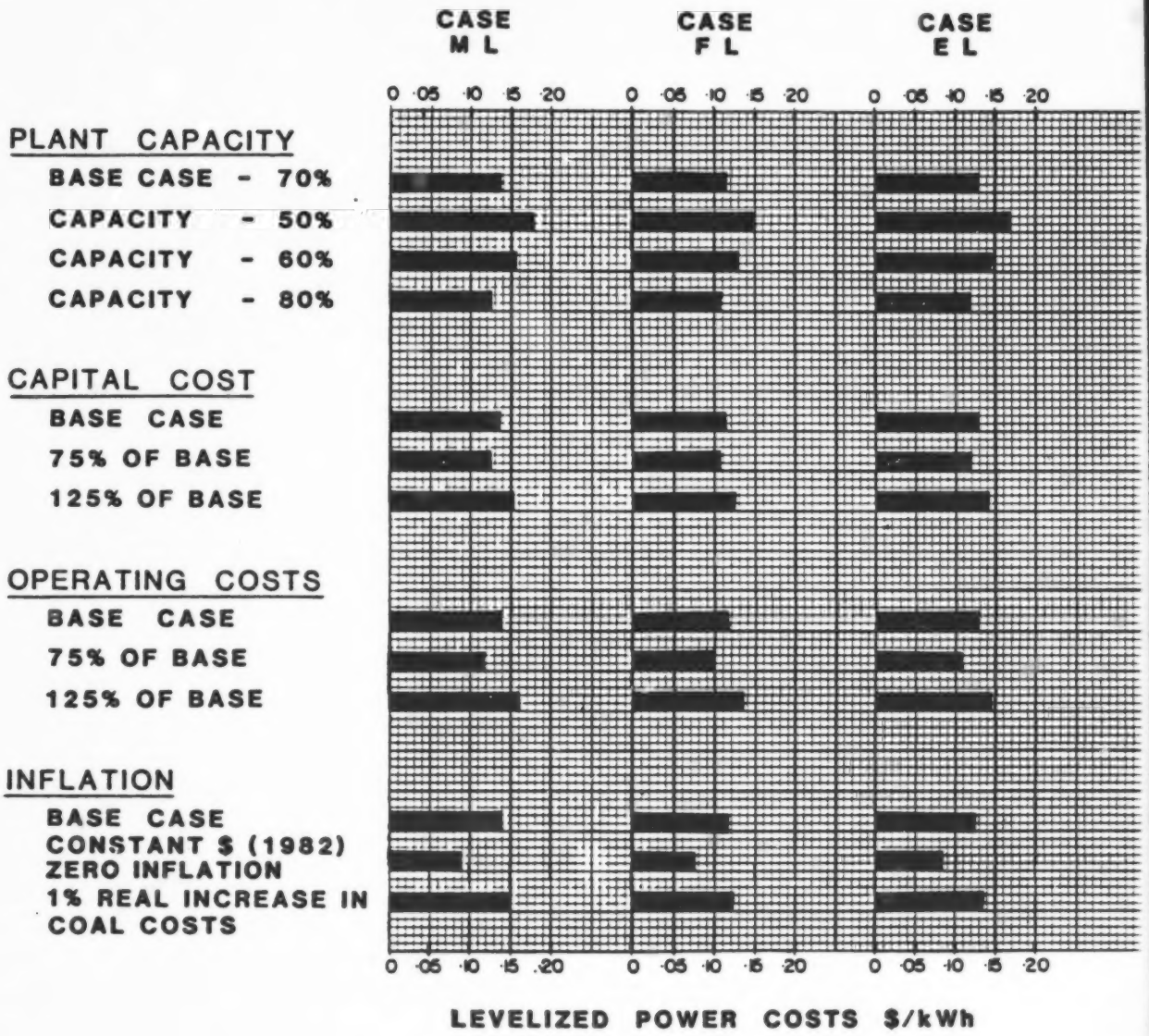
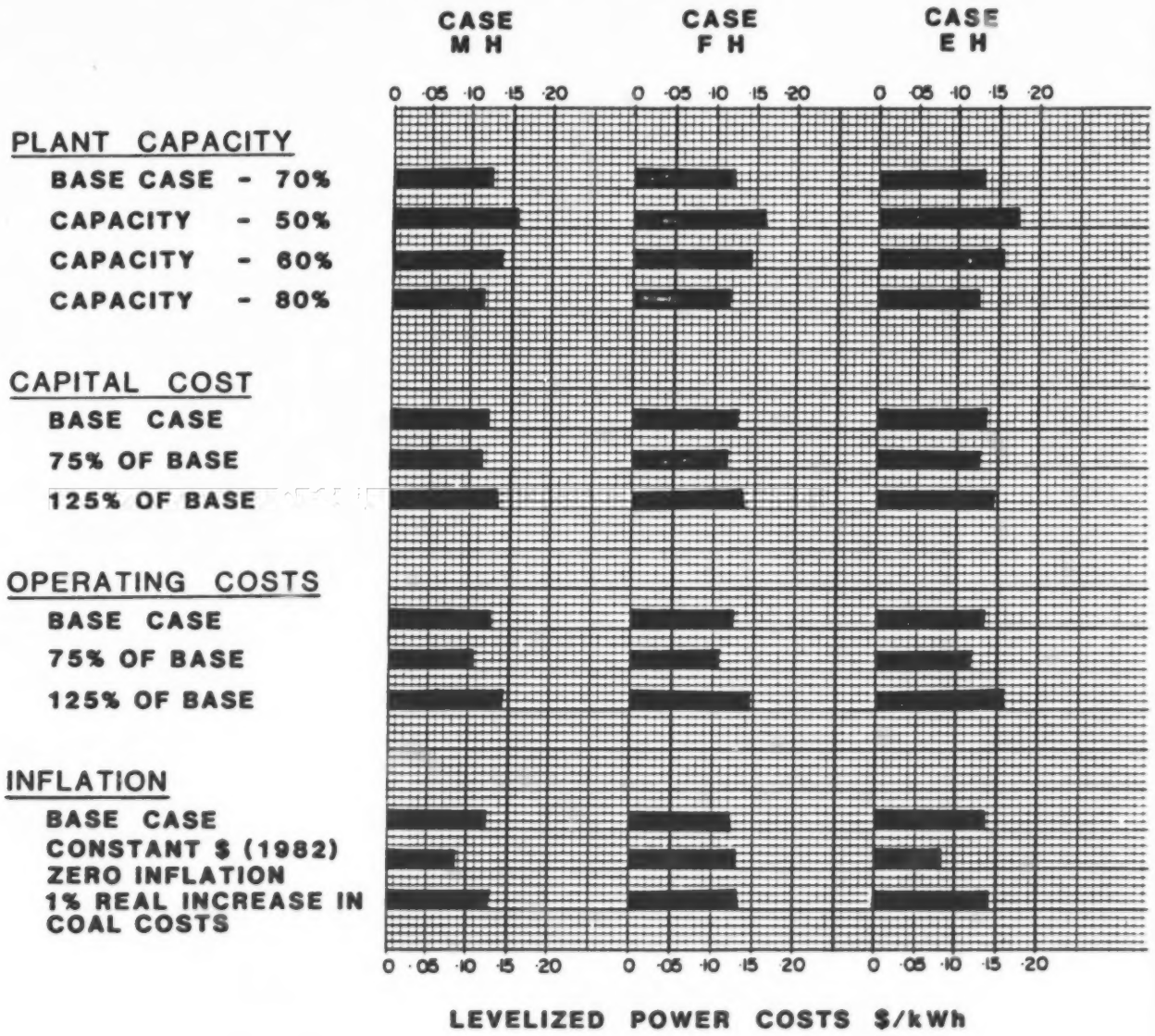


FIGURE No 8.2.3-2
GRAPHICAL SUMMARY OF SENSITIVITY ANALYSIS
HIGH TEMPERATURE PURIFICATION OPTIONS



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